PB-231 757

AIR POLLUTION CONTROL TECHNOLOGY AND COSTS IN SEVEN SELECTED AREAS

Industrial Gas Cleaning Institute, Incorporated

PREPARED FOR
Environmental Protection Agency

December 1973

DISTRIBUTED BY:
NTIS
National Technical Information Service
U. S. DEPARTMENT OF COMMERCE

BIBLIOGRAPHIC DATA SHEET

1. Report No. EPA-450/3-73-010

2.

PB 231 757

4. Title and Subtitle

Air Pollution Control Technology and Costs in Seven Selected Areas

5. Report Date Dec., 1973 (Date of issue)

6.

7. Author(s)

8. Performing Organization Rept. No. 47.173

9. Performing Organization Name and Address

Industrial Gas Cleaning Institute
P.O. Box 1333
Stamford, Connecticut 06904

10. Project/Task/Work Unit No.

11. Contract/Grant No.

68-02-0289

12. Sponsoring Organization Name and Address

Environmental Protection Agency
Durham Contract Operations
Research Triangle Park
North Carolina, 27711

13. Type of Report & Period Covered

Final

14.

15. Supplementary Notes

16. Abstracts

Under this contract, the Industrial Gas Cleaning Institute collected and formalized date on air pollution abatement in eight selected industrial areas. Those eight areas were: Phosphate Fertilizer Manufacture, Feed and Grain Milling, Soap and Detergent Manufacture, Paint and Varnish Production, The Graphic Arts Industry, Lime Kiln Operation, and Gray Iron Foundry Cupola Operation. For each area studied, costs of conventionally applied pollution control systems are presented for a range of plant sizes and control efficiencies.

17. Key Words and Document Analysis. 17a. Descriptors

Air Pollution
Performance
Air Pollution Control Equipment
Cost Estimates
Cost Engineering
Expenses
Electrostatic Precipitators
Scrubbers
Incinerators
Fabric Filters

17b. Identifiers/Open-Ended Terms

Air Pollution Control
Graphic Arts
Phosphate Fertilizer
Paint and Varnish
Lime Kilns
Feed and Grain
Soap and Detergent
Gray Iron Foundries

Reproduced by
NATIONAL TECHNICAL
INFORMATION SERVICE
U S Department of Commerce
Springfield VA 22151

17c. COSATI Field/Group 13.b

18. Availability Statement

Release Unlimited

19. Security Class (This Report) UNCLASSIFIED

20. Security Class (This Page) UNCLASSIFIED

21. No. of Pages 708

22. Price 13.00 / 1.45

FORM NTIS-35 (REV. 3-72)

USCOMM-DC 14952-P72

EPA-450/3-73-010

AIR POLLUTION CONTROL TECHNOLOGY AND COSTS IN SEVEN SELECTED AREAS

by

Industrial Gas Cleaning Institute
P.O. Box 1333
Stamford, Connecticut 06904

Contract No. 68-02-0289

EPA Project Officer: Paul A. Boys

Prepared for

ENVIRONMENTAL PROTECTION AGENCY
Office of Air and Water Programs
Office of Air Quality Planning and Standards
Research Triangle Park, N.C. 27711

December 1973

1a

This report is issued by the Environmental Protection Agency to report technical data of interest to a limited number of readers. Copies are available free of charge to Federal employees, current contractors and grantees, and nonprofit organizations - as supplies permit - from the Air Pollution Technical Information Center, Environmental Protection Agency, Research Triangle Park, North Carolina 27711, or from the National Technical Information Service, 5285 Port Royal Road, Springfield, Virginia 22151.

This report was furnished to the Environmental Protection Agency by Industrial Gas Cleaning Institute, Stamford, Connecticut, in fulfillment of Contract No. 68-02-0289. The contents of this report are reproduced herein as received from the contractor. The opinions, findings, and conclusions expressed are those of the author and not necessarily those of the Environmental Protection Agency.

Publication No. EPA-450/3-73-010

CONTRACT NUMBER 68-02-0289

AIR POLLUTION CONTROL
TECHNOLOGY AND COSTS
IN SEVEN SELECTED AREAS

FINAL REPORT

(Submitted December 15, 1973)

by

L. C. Hardison, Coordinating Engineer

for

INDUSTRIAL GAS CLEANING INSTITUTE

Box 1333, Stamford, Connecticut 06904

Prepared For

THE ENVIRONMENTAL PROTECTION AGENCY

Durham, North Carolina 27701

INDUSTRIAL GAS CLEANING INSTITUTE, INC.

Box 1333
Stamford, Connecticut 06904

MEMBERS

- AIR CORRECTION DIVISION
  - UNIVERSAL OIL PRODUCTS CO.
- AMERICAN AIR FILTER COMPANY, INC.
- AMERICAN STANDARD, INC.
  - INDUSTRIAL PRODUCTS DIVISION
- BELCO POLLUTION CONTROL CORP.
- BUFFALO FORGE COMPANY
- THE CARBORUNDUM COMPANY
  - POLLUTION CONTROL DIVISION
- C E AIR PREHEATER
- THE CEILCOTE COMPANY, INC.
- CHEMICAL CONSTRUCTION CORP.
  - POLLUTION CONTROL DIVISION
- THE DUCON COMPANY, INC.
  - SUBS. OF U.S. FILTER CORP.
- DUSTEX DIVISION
  - AMERICAN PRECISION INDUSTRIES, INC.
- ENVIRONEERING, INC.
  - SUBS. OF THE RILEY COMPANY
- ENVIROTECH-CORPORATION
  - ARCO DIVISION
  - BUELL DIVISION
  - NORBLO DIVISION
- FISHER-KLOSTERMAN, INC.
- FULLER COMPANY, DRACCO PRODUCTS
  - SUBS. OF GENERAL AMERICAN TRANSPORTATION CORP.
- GALLAGHER-KAISER CORPORATION
- INDUSTRIAL CLEAN AIR, INC.
- JOHNS MANVILLE CORP.,
  - ENVIRONMENTAL SYSTEMS GROUP
- THE KIRK & BLUM MANUFACTURING CO.
- THE KOPPERS COMPANY, INC.
  - ENVIRONMENTAL SYSTEMS DIVISION
- MATHEY-BISHOP, INC.
- MIKROPUL, DIV. OF U.S. FILTER CORP.
- PEABODY ENGINEERING CORPORATION
- POLLUTION CONTROL-WALTHER, INC.
- PRECIPITATION ASSOCIATES OF AMERICA, INC.
- RESEARCH-COTTRELL, INC.
- THE TORIT CORPORATION
- WESTERN PRECIPITATION DIVISION
  - JOY MANUFACTURING COMPANY
- WHEELABRATOR-FRYE, INC.
  - AIR POLLUTION CONTROL DIVISION
- ZURN INDUSTRIES, INC.
  - AIR SYSTEMS DIVISION

STATEMENT OF PURPOSES

The Industrial Gas Cleaning Institute, incorporated in 1960 in the State of New York, was founded to further the interests of manufacturers of air pollution control equipment, by

encouraging the general improvement of engineering and technical standards in the manufacture, installation, operation, and performance of equipment

disseminating information on air pollution; the effect of industrial gas cleaning on public health; and general economic, social, scientific, technical, and governmental matters affecting the industry, together with the views of the members thereon; and

promoting the industry through desirable advertising and publicity.

ACKNOWLEDGEMENT

The efforts of the IGCI Engineering Standards Committee in the preparation, editing and reviewing of this Report are gratefully acknowledged:

Harry Krockta, Chairman, The Ducon Company, Inc.
G. L. Brewer, Air Correction Division, UOP
C. A. Gallaer, Buell Division of Envirotech Corp.
N. D. Phillips, Fuller Co., Dracco Products
E. P. Stastny, Koppers Company, Inc., Environmental Systems Division

TABLE OF CONTENTS

Page No.

I. INTRODUCTION 1

II. TECHNICAL DATA 2
- A. General Description
  - 1. Format
  - 2. Selection of Applicable Equipment Types
  - 3. Basis for Preparing Specifications and Bid Prices
  - 4. Presentation of Data

- B. Process Description and Costs
  - 1. Phosphate Industry 11
    - a. Wet Phosphoric Acid (WPPA)
    - b. Super Phosphoric Acid (SPA)
    - c. Granular Triple Superphosphate (GTSP)
    - d. Diammonium Phosphate (DAP)
    - e. GTSP Storage
  - 2. Feed and Grain Industry 193
    - a. Feed Mills
    - b. Flour Mills
    - c. Feed Flash Dryers
  - 3. Paint and Varnish Industry 249
    - a. Open Kettles
    - b. Closed Reactors
  - 4. Graphic Arts 331
    - a. Web Offset Printing
    - b. Metal Decorating
    - c. Gravure Printing
  - 5. Soap and Detergent 429
    - a. Spray Dryers
    - b. Dry Product Handling
  - 6. Lime Kilns 481
    - a. Vertical Rock Kilns
    - b. Rotary Sludge Kilns
  - 7. Gray Iron Foundaries 561
    - a. Cupolas

TABLE OF CONTENTS
(continued)

Page No.

C. Additional Cost Data

1. Discussion of Cost Basis 605
2. Operating Cost at Various Utility Cost Levels 607
3. Derived Cost Indices 610
4. Generalized Cost Data 677

List of Figures

List of Tables

List of Appencides

LIST OF FIGURES

Page No.

Figure 1 Outline Map of U.S. Indicating Phosphate Producing Areas 13
Figure 2 Block Diagram of Phosphate Fertilizer Manufacturing Processes 15
Figure 3 Simplified Flow Scheme for Strong Acid Process 21
Figure 4 Schematic Flow Diagram for Wet Phosphoric Acid Process 23
Figure 5 Contribution of HF and SiF_4 to Vapor Phase Fluorine Content at 50°C 44
Figure 6 Concentration of Fluorine in Vapor Over H_2SiF_6 Solutions 45
Figure 7 Concentration of Fluorine in Vapor Over H_2SiF_6 Solutions Vs. Temperature 46
Figure 8 Comparison of Scrubber Types 49
Figure 9 Sketch of Cross-flow Packed Scrubber 50
Figure 10 Combination Scrubber for WPPA Process 51
Figure 11 Capital Costs for Wet Scrubbers for WPPA Process Plants 61
Figure 12 Confidence Limits for Capital Cost of Scrubbers for WPPA Process Plants 63
Figure 13 Confidence Limits for Capital Cost of Scrubbers for WPPA Process Plants 65
Figure 14 Annual Cost for Wet Scrubbers for WPPA Process Plants 67
Figure 15 Schematic Flow Diagram for Vacuum Flash Superphosphoric Acid Process 69
Figure 16 Schematic Representation of Polyphosphoric Acid Forms 75
Figure 17 Plot of Barometric Condenser Temperature Vs. Cooling Water/ Steam Ratio 78
Figure 18 Schematic Drawing of Water-Induced Venturi Scrubber 81
Figure 19 Schematic Drawing of Venturi Scrubber with Packed Entrainment Separater 83
Figure 20 Capital Cost for Water Induced Venturi Scrubbers for SPA Process Plants 89
Figure 21 Confidence Limits for Capital Cost of Water Induced Venturi Scrubbers for SPA Process Plants 91
Figure 22 Confidence Limits for Installed Cost of Water Induced Venturi Scrubbers for SPA Process Plants 93
Figure 23 Annual Cost for Water Induced Venturi Scrubbers for SPA Process Plants 95
Figure 24 Capital Cost for Venturi Scrubbers with Packed Section for SPA Process Plants 97
Figure 25 Confidence Limits for Capital Cost of Venturi Scrubbers with Packed Section for SPA Process Plants 99

Page No.

Figure 26 Confidence Limits for Installed Cost of Venturi Scrubber with Packed Section for SPA Process Plants 101
Figure 27 Annual Cost for Venturi Scrubbers with Packed Section for SPA Process Plants 103
Figure 28 Flow Diagram of Granular Triple Superphosphate Plant 105
Figure 29 Flow Scheme of Phosphate Rock Milling Operation 109
Figure 30 Capital Cost of Wet Scrubbers for GTSP Process Plants 121
Figure 31 Confidence Limits for Capital Cost of Wet Scrubbers for GTSP Process Plants 123
Figure 32 Confidence Limits for Capital Cost of Installed Wet Scrubbers for GTSP Process Plants 125
Figure 33 Annual Cost of Wet Scrubbers for GTSP Process Plants 127
Figure 34 Schematic Flow Drawing for Diammonium Phosphate Process 129
Figure 35 Schematic Drawing of Scrubbers Used in DAP Tail Gas Treating . . . 142
Figure 36 Acceptable Scrubber Combinations for DAP Process Plants 144
Figure 37 Capital Cost of Two Stage Cyclonic Scrubbers for DAP Process Plants 151
Figure 38 Confidence Limits for Capital Cost of Two Stage Cyclonic Scrubbers for DAP Process Plants 153
Figure 39 Confidence Limits for Capital Cost of Installed Two Stage Cyclonic Scrubbers for DAP Process Plants 155
Figure 40 Annual Cost of Two Stage Cyclonic Scrubbers for DAP Process Plants 157
Figure 41 Capital Cost of Venturi Cyclone Scrubbers for DAP Process Plants . . 159
Figure 42 Confidence Limits for Capital Cost of Venturi Cyclone Scrubbers for DAP Process Plants 161
Figure 43 Confidence Limits for Capital Cost of Installed Venturi Cyclone Scrubbers for DAP Process Plants 163
Figure 44 Annual Cost of Venturi Cyclone Scrubbers for DAP Process Plants . 165
Figure 45 Capital Cost for Packed Cross-flow Scrubbers for DAP Process Plants 169
Figure 46 Confidence Limits for Capital Cost of Packed Cross-flow Scrubbers for DAP Process Plants 171
Figure 47 Confidence Limits for Installed Cost of Packed Cross-flow Scrubbers for DAP Process Plants 173
Figure 48 Annual Cost of Packed Cross-Flow Scrubbers for DAP Process Plants. 177
Figure 49 Sketch of GTSP Storage Building with 4,000,000 ft^3 Interior 180
Figure 50 GTSP Storage Building with Cyclonic Scrubbers 183

Page No.

Figure 51 Simplified Flow Diagram of a Feed Mill 199
Figure 52 Simplified Flow Diagram of a Modern Flour Mill 202
Figure 53 Flash Drying Combined with Disintegration 207
Figure 54 Capital Costs for Fabric Filters for Flour Milling 219
Figure 55 Annual Costs for Fabric Filters for Flour Milling 221
Figure 56 Confidence Limits for Capital Cost of Fabric Filters for Flour Milling 223
Figure 57 Capital Costs for Fabric Filters for Feed Grinding 227
Figure 58 Annual Costs for Fabric Filters for Feed Grinding 229
Figure 59 Confidence Limits for Collector Only Cost of Fabric Filters for Feed Grinding 231
Figure 60 Capital Costs for Wet Scrubbers for Feed Flash Dryers (High Efficiency) 235
Figure 61 Annual Costs for Wet Scrubbers for Feed Flash Dryers (High Efficiency) 237
Figure 62 Confidence Limits for Capital Cost of Wet Scrubbers for Feed Flash Dryers (High Efficiency) 239
Figure 63 Capital Costs for Thermal Incinerators for Feed Flash Dryers 243
Figure 64 Annual Costs for Thermal Incinerators for Feed Flash Dryers 245
Figure 65 Confidence Limits for Capital Cost of Thermal Incinerators for Feed Flash Dryers 247
Figure 66 Paint Manufacturing Using Sand Mill for Grinding Operation 251
Figure 67 Materials Flow Sheet for Paint Manufacturing 252
Figure 68 Typical Varnish Cooking Room 255
Figure 69 Modern Resin Production System 263
Figure 70 Hydrocarbon Emission for Dowtherm Kettle 270
Figure 71 Hydrocarbon Emission for Paint Kettle 271
Figure 72 Kettle Hydrocarbon Emission – 2 Batches Monitored . 272
Figure 73 Pigment Emission Control System 274
Figure 74 Thermal Combustion System for Resin Reactor or Closed Kettle 276
Figure 75 Schematic Diagram of a Catalytic Combustion System for Varnish Kettles 278
Figure 76 Capital Costs for Thermal Incinerators for the Paint and Varnish Industry (Without Heat Exchange) 306

Page No.

Figure 77 Capital Costs for Catalytic Incinerators for the Paint and Varnish Industry (Without Heat Exchange) 307
Figure 78 Direct Annual Costs for Thermal and Catalytic Incinerators for the Paint and Varnish Industry (Without Heat Exchange) 308
Figure 79 Total Annual Costs for Thermal and Catalytic Incinerators for the Paint and Varnish Industry (Without Heat Exchange) 309
Figure 80 Capital Costs for Thermal Incinerators for the Paint and Varnish Industry (With Heat Exchange) 322
Figure 81 Capital Costs for Catalytic Incinerators for the Paint and Varnish Industry (With Heat Exchange) .. 323
Figure 82 Direct Annual Costs for Thermal and Catalytic Incinerators for the Paint and Varnish Industry (With Heat Exchange) 324
Figure 83 Total Annual Costs for Thermal and Catalytic Incinerators for the Paint and Varnish Industry (With Heat Exchange) 325
Figure 84 Total Installed Costs for Thermal and Catalytic Incinerators for the Paint and Varnish Industry 327
Figure 85 Web Offset, Publication 342
Figure 86 Emission Rates from Web Offset and Web Letterpress Employing Heatset Inks 345
Figure 87 Web Letterpress, Publication 346
Figure 88 Web Letterpress, Newspaper 348
Figure 89 Rotogravure Printing Operation 349
Figure 90 Emission Rates from a Typical Rotogravure Printing Operation 351
Figure 91 Metal Decorating Coating Operation 352
Figure 92 Metal Decorating Printing and Varnish Overcoating ... 353
Figure 93 Products Produced at Various Oxidation Stages During Combustion 357
Figure 94 Flow Diagram for Thermal Combustion Including Possibilities for Heat Recovery 358
Figure 95 Flow Diagram for Catalytic Combustion Including Possibilities for Heat Recovery 359
Figure 96 Flow Diagram of Adsorption Process 360
Figure 97 Schematic Flow Diagram of Thermal and Catalytic Combustion Without Heat Exchange for Web Offset and Metal Decorating 365
Figure 98 Capital Costs for Thermal Incinerators for Web Offset Printing (Without Heat Exchange) 369

Page No.

Figure 99 Annual Costs for Thermal Incinerators for Web Offset Printing (Without Heat Exchange) 371
Figure 100 Schematic Flow Diagram of Thermal and Catalytic Combustion Without Heat Exchange for Web Offset and Metal Decorating 373
Figure 101 Capital Costs for Catalytic Incinerators for Web Offset Printing (Without Heat Exchange) 377
Figure 102 Annual Costs for Catalytic Incinerators for Web Offset Printing (Without Heat Exchange) 379
Figure 103 Schematic Flow Diagram of Thermal and Catalytic Combustion Without Heat Exchange for Web Offset and Metal Decorating 381
Figure 104 Capital Costs for Thermal Incinerators for Metal Decorating (Without Heat Exchange) 385
Figure 105 Annual Costs for Thermal Incinerators for Metal Decorating (Without Heat Exchange) 387
Figure 106 Schematic Flow Diagram of Thermal and Catalytic Combustion With Heat Exchange for Web Offset and Metal Decorating 389
Figure 107 Capital Costs for Thermal Incinerators for Metal Decorating (With Heat Exchange) 393
Figure 108 Annual Costs for Thermal Incinerators for Metal Decorating (With Heat Exchange) 395
Figure 109 Schematic Flow Diagram of Thermal and Catalytic Combustion Without Heat Exchange for Web Offset and Metal Decorating 397
Figure 110 Capital Costs for Catalytic Incinerators for Metal Decorating (Without Heat Exchange) 401
Figure 111 Annual Costs for Catalytic Incinerators for Metal Decorating (Without Heat Exchange) 403
Figure 112 Schematic Flow Diagram of Thermal and Catalytic Combustion With Heat Exchange for Web Offset and Metal Decorating 405
Figure 113 Capital Costs for Catalytic Incinerators for Metal Decorating (With Heat Exchange) 409
Figure 114 Annual Costs for Catalytic Incinerators for Metal Decorating (With Heat Exchange) 411
Figure 115 Schematic Flow Diagram of Thermal Combustion With Heat Exchange for Gravure Printing 413
Figure 116 Capital Costs for Thermal Incinerators for Gravure Printing (With Heat Exchange) 417
Figure 117 Annual Costs for Thermal Incinerators for Gravure Printing (With Heat Exchange) 419

Page No.

Figure 118 Capital Costs for Carbon Adsorption for Gravure Printing 423
Figure 119 Annual Costs for Carbon Adsorption for Gravure Printing 425
Figure 120 Confidence Limits for Capital Cost of Carbon Adsorption for Gravure Printing 427
Figure 121 Over-All Routes to Linear Alkylate Sulfonates (LAS), Alpha-Olefin Sulfonates (AOS) and Alcohol Sulfates 432
Figure 122 Low- and High-Active Sulfonation Using Oleum 433
Figure 123 Alpha-Olefin Sulfonation Flowsheet 434
Figure 124 Typical Spray Drying Plant 437
Figure 125 Continuous Process for Soap Manufacture 440
Figure 126 Capital Costs for Wet Scrubbers for Soap and Detergent Spray Drying (Medium Efficiency) 456
Figure 127 Annual Costs for Wet Scrubbers for Soap and Detergent Spray Drying (Medium Efficiency) 457
Figure 128 Capital Costs for Wet Scrubbers for Soap and Detergent Spray Drying (High Efficiency) 458
Figure 129 Annual Costs for Wet Scrubbers for Soap and Detergent Spray Drying (High Efficiency) 459
Figure 130 Confidence Limits for Capital Cost of Wet Scrubbers for Soap and Detergent Spray Drying (Medium Efficiency) 461
Figure 131 Confidence Limits for Capital Cost of Wet Scrubbers for Soap and Detergent Spray Drying (High Efficiency) . 463
Figure 132 Capital Costs for Fabric Filters for Soap and Detergent Spray Drying 467
Figure 133 Annual Costs for Fabric Filters for Soap and Detergent Spray Drying 469
Figure 134 Confidence Limits for Collector Only Cost of Fabric Filters for Soap and Detergent Spray Drying 471
Figure 135 Capital Costs for Fabric Filters for Soap and Detergent Product Handling 475
Figure 136 Annual Costs for Fabric Filters for Soap and Detergent Product Handling 477
Figure 137 Confidence Limits for Collector Only Cost of Fabric Filters for Soap and Detergent Product Handling 479
Figure 138 Simplified Flow Sheet of a Typical Lime Manufacturing Operation 482
Figure 139 Vertical Lime Kiln 484
Figure 140 Simplified Kraft Mill Flow Diagram 491
Figure 141 Kraft Pulping Recausticizing Flow Sheet 492
Figure 142 Cyclonic Gas Scrubber 508

Page No.

Figure 143 Variable Orifice Scrubber 509
Figure 144 Lime Kiln Venturi Scrubber System 510
Figure 145 Capital Costs for Wet Scrubbers for Vertical Lime Rock Kilns (Medium Efficiency) 516
Figure 146 Annual Costs for Wet Scrubbers for Vertical Lime Rock Kilns (Medium Efficiency) 517
Figure 147 Capital Costs for Wet Scrubbers for Vertical Lime Rock Kilns (High Efficiency) 518
Figure 148 Annual Costs for Wet Scrubbers for Vertical Lime Rock Kilns (High Efficiency) 519
Figure 149 Confidence Limits for Capital Cost of Wet Scrubbers for Vertical Lime Rock Kilns (High Efficiency) 521
Figure 150 Capital Costs for Fabric Filters for Vertical Lime Rock Kilns 525
Figure 151 Annual Costs for Fabric Filters for Vertical Lime Rock Kilns 527
Figure 152 Capital Costs for Electrostatic Precipitators for Vertical Lime Rock Kilns (Medium Efficiency) 532
Figure 153 Annual Costs for Electrostatic Precipitators for Vertical Lime Rock Kilns (Medium Efficiency) 533
Figure 154 Capital Costs for Electrostatic Precipitators for Vertical Lime Rock Kilns (High Efficiency) 534
Figure 155 Annual Costs for Electrostatic Precipitators for Vertical Lime Rock Kilns (High Efficiency) 535
Figure 156 Confidence Limits for Capital Cost of Electrostatic Precipitators for Vertical Lime Rock Kilns (High Efficiency) 537
Figure 157 Capital Costs for Wet Scrubbers for Rotary Lime Sludge Kilns (Medium Efficiency) 542
Figure 158 Annual Costs for Wet Scrubbers for Rotary Lime Sludge Kilns (Medium Efficiency) 543
Figure 159 Capital Costs for Wet Scrubbers for Rotary Lime Sludge Kilns (High Efficiency) 544
Figure 160 Annual Costs for Wet Scrubbers for Rotary Lime Sludge Kilns (High Efficiency) 545
Figure 161 Confidence Limits for Capital Cost of Wet Scrubbers for Rotary Lime Sludge Kilns (High Efficiency) 547
Figure 162 Capital Costs for Electrostatic Precipitators for Rotary Lime Sludge Kilns (Medium Efficiency) 552
Figure 163 Annual Costs for Electrostatic Precipitators for Rotary Lime Sludge Kilns (Medium Efficiency) 553
Figure 164 Capital Costs for Electrostatic Precipitators for Rotary Lime Sludge Kilns (High Efficiency) 554

Page No.

Figure 165 Annual Costs for Electrostatic Precipitators for Rotary Lime Sludge Kilns (High Efficiency) 555
Figure 166 Confidence Limits for Capital Cost of Electrostatic Precipitators for Rotary Lime Sludge Kilns (High Efficiency) 557
Figure 167 Distribution of Iron Foundry Sizes (1969) 562
Figure 168 Schematic Flow Diagram of Cast Iron Production 566
Figure 169 Flow Diagram of Major Iron Foundry Operations 567
Figure 170 Trends in Types of Iron Foundry Furnaces 569
Figure 171 Typical Gray Iron Cupola 571
Figure 172 Capital Costs for Fabric Filters for Gray Iron Foundry Cupolas 587
Figure 173 Annual Costs for Fabric Filters for Gray Iron Foundry Cupolas 589
Figure 174 Confidence Limits for Capital Cost of Fabric Filters for Gray Iron Foundry Cupolas 591
Figure 175 Capital Costs for Wet Scrubbers for Gray Iron Foundry Cupolas (Medium Efficiency) 598
Figure 176 Annual Costs for Wet Scrubbers for Gray Iron Foundry Cupolas (Medium Efficiency) 599
Figure 177 Capital Costs for Wet Scrubbers for Gray Iron Foundry Cupolas (High Efficiency) 600
Figure 178 Annual Costs for Wet Scrubbers for Gray Iron Foundry Cupolas (High Efficiency) 601
Figure 179 Confidence Limits for Capital Cost of Wet Scrubbers for Gray Iron Foundry Cupolas (High Efficiency) ... 603

LIST OF TABLES

Page No.

Table 1 Abatement Equipment Types Selected for the Seven Process Areas . . 5
Table 2 Process Weight Table for Medium Efficiency for Particulate Collection Cases Only 8
Table 3 Performance Levels Selected for Phase II and III Areas 9
Table 4 Typical Composition and Particle Size of Commercial Grades of Florida Phosphate Rock 28
Table 5 Analyses of Pebble Phosphate Rock 29
Table 6 Feed to WPPA Process Before and After Calcination 30
Table 7 Wet Phosphoric Acid Product Breakdown 32
Table 8 Distribution of Fluorine in WPPA Plant Discharge Streams 33
Table 9 Florida Fluoride Emission Law 36
Table 10 Estimated Abatement Requirements Under Florida Law 39
Table 11 Calculated Concentrations of SiF_4 and HF at 50°C 41
Table 12 Calculated Concentrations of Fluorine in Vapor Phase at 60°C and 70°C 43
Table 13 Concentration Limits at Scrubber Outlet 53
Table 14 NTU Required to Reach 2.15 ppm F Discharge Concentration 55
Table 15 Process Description for Wet Process Phosphoric Acid Plant Cross-flow Scrubber 57
Table 16 Operating Conditions for Wet Process Phosphoric Acid Plant Cross-flow Scrubber 58
Table 17 Estimated Capital Cost Data (Costs in Dollars) for Wet Scrubbers for WPPA Process Plants 60
Table 18 Confidence Limits for Capital Cost of Scrubbers for WPPA Process Plants 62
Table 19 Confidence Limits for Capital Cost of Installed Scrubbers for WPPA Process Plants 64
Table 20 Annual Operating Cost Data (Costs in $/Year) for Wet Scrubbers for WPPA Process Plants 66
Table 21 Forms of Phosphoric Acid 74
Table 22 Process Description for Wet Scrubber for SPA Process Specification 84
Table 23 Operating Conditions for Wet Scrubber 86
Table 24 Capital Cost Data (Costs in Dollars) for Water Induced Venturi Scrubbers for SPA Process Plants 88
Table 25 Confidence Limits for Capital Cost of Water Induced Venturi Scrubbers for SPA Process Plants 90

Page No.

Table 26 Confidence Limits for Installed Cost of Water Induced Venturi Scrubbers for SPA Process Plants 92

Table 27 Annual Operating Cost Data (Costs in $/Year) for Water Induced Venturi Scrubbers for SPA Process Plants 94

Table 28 Capital Cost Data (Costs in Dollars) for Venturi Scrubber with Packed Section for SPA Process Plants 96

Table 29 Confidence Limits for Capital Cost of Venturi Scrubbers with Packed Section for SPA Process Plants 98

Table 30 Confidence Limits for Installed Cost of Venturi Scrubbers with Packed Section for SPA Process Plants 100

Table 31 Annual Operating Cost Data (Costs in $/Year) for Venturi Scrubber with Packed Section for SPA Process Plants 102

Table 32 Material Balance of GTSP Process for 250 Ton/Day 111

Table 33 Materials Evolved as Gases from Triple Superphosphate Manufacture for 250 Ton/Day Process 113

Table 34 Process Description for Granular Triple Superphosphate Specification 117

Table 35 Operating Conditions for Granular Triple Superphosphate Wet Scrubber Specification 119

Table 36 Estimated Capital Cost Data (Costs in Dollars) for Wet Scrubbers for GTSP Process Plants 120

Table 37 Confidence Limits for Capital Cost of Wet Scrubbers for GTSP Process Plants 122

Table 38 Confidence Limits for Installed Cost of Wet Scrubbers for GTSP Process Plants 124

Table 39 Annual Operating Cost Data (Costs in $/Year) for Wet Scrubbers for GTSP Process Plants 126

Table 40 Approximate Material Balance of 1,000 Ton/Day DAP Process 134

Table 41 Combined Vents from Reactor and Granulator 136

Table 42 Rough Heat Balance for DAP Dryer for 1,000 Ton/Day Process 138

Table 43 Process Description for Wet Scrubbers for DAP Process 145

Table 44 Operating Conditions for Primary Scrubbers for DAP Process Dryer Vents 147

Table 45 Operating Conditions for Secondary Scrubbers for DAP Process Dryer Vents 148

Table 46 Estimated Capital Cost Data (Costs in Dollars) for Two Stage Cyclonic Scrubbers for DAP Process Plants 150

Table 47 Confidence Limits for Capital Cost of Two Stage Cyclonic Scrubbers for DAP Process Plants 152

Table 48 Confidence Limits for Installed Cost of Two Stage Cyclonic Scrubbers for DAP Process Plants 154

Page No.

Table 49 Annual Operating Cost Data (Costs in $/Year) for Two Stage Cyclonic Scrubbers for DAP Process Plants 156
Table 50 Estimated Capital Cost Data (Costs in Dollars) for Venturi Cyclonic Scrubbers for DAP Process Plants 158
Table 51 Confidence Limits for Capital Cost of Venturi Cyclonic Scrubbers for DAP Process Plants 160
Table 52 Confidence Limits for Installed Cost of Venturi Cyclonic Scrubbers for DAP Process Plants 162
Table 53 Annual Operating Cost Data (Costs in $/Year) for Venturi Cyclonic Scrubbers for DAP Process Plants 164
Table 54 Estimated Capital Cost Data (Costs in Dollars) for Packed Cross-flow Scrubbers for DAP Process Plants 167
Table 55 Confidence Limits for Capital Cost of Packed Cross-flow Scrubbers for DAP Process Plants 170
Table 56 Confidence Limits for Installed Cost of Packed Cross-flow Scrubbers for DAP Process Plants 172
Table 57 Annual Operating Cost Data (Costs in $/Year) for Packed Cross-flow Scrubbers for DAP Process Plants 175
Table 58 Process Description for GTSP Storage Building Scrubber 185
Table 59 Operating Conditions for Scrubber Specification for GTSP Storage Building Vent 186
Table 60 Estimated Capital Cost Data (Costs in Dollars) for Cyclonic Scrubbers for GTSP Storage Vents 187
Table 61 Confidence Limits for Capital Cost of Cyclonic Scrubbers for GTSP Storage 188
Table 62 Confidence Limits for Installed Cost of Cyclonic Scrubbers for GTSP Storage 189
Table 63 Annual Operating Cost Data (Costs in $/Year) for Cyclonic Scrubbers for GTSP Storage Vents 190
Table 64 Composition of Typical Feedstuffs Used in Manufacture of Feed ... 194
Table 65 Rations and Supplements for Brood Sows 196
Table 66 Beef Cattle Feed Supplements 197
Table 67 Sources of Air Pollution in Feed and Grain Industry 209
Table 68 Particle Size Analysis of the Primary Cyclone Catch and the Secondary Fabric Filter Catch of Dust from a Railroad Receiving Hopper Hood Controlling the Unloading of a Boxcar of Feed-Type Barley 210
Table 69 Dust Losses from Cyclones 212
Table 70 Fabric Filter Process Description for Flour Milling Specification ... 216
Table 71 Fabric Filter Operating Conditions for Flour Milling Specification ... 217
Table 72 Estimated Capital Cost Data for Fabric Filters for Flour Milling 218

Page No.

Table 73 Annual Operating Cost Data for Fabric Filters for Flour Milling 220
Table 74 Confidence Limits for Capital Cost of Fabric Filters for Flour Milling 222
Table 75 Fabric Filter Process Description for Feed Grinding Specification 224
Table 76 Fabric Filter Operating Conditions for Feed Grinding Specification 225
Table 77 Estimated Capital Cost Data for Fabric Filters for Feed Grinding 226
Table 78 Annual Operating Cost Data for Fabric Filters for Feed Grinding 228
Table 79 Confidence Limits for Collector Only Cost of Fabric Filters for Feed Grinding 230
Table 80 Wet Scrubber Process Description for Feed Flash Dryer Specification 232
Table 81 Wet Scrubber Operating Conditions for Feed Flash Dryer Specification 233
Table 82 Estimated Capital Cost Data for Wet Scrubbers for Feed Flash Dryers 234
Table 83 Annual Operating Cost Data for Wet Scrubbers for Feed Flash Dryers 236
Table 84 Confidence Limits for Capital Cost of Wet Scrubbers for Feed Flash Dryers (High Efficiency) 238
Table 85 Thermal Incinerator Process Description for Feed Flash Dryer Specification 240
Table 86 Thermal Incinerator Operating Conditions for Feed Flash Dryer Specification 241
Table 87 Estimated Capital Cost Data for Thermal Incinerators for Feed Flash Dryers 242
Table 88 Annual Operating Cost Data for Thermal Incinerators for Feed Flash Dryers 244
Table 89 Confidence Limits for Capital Cost of Thermal Incinerators for Feed Flash Dryers 246
Table 90 Varnish Raw Material 254
Table 91 Composition of Oil and Varnish Fumes 267
Table 92 Emission Data Summary 269
Table 93 Thermal Incinerator Process Description for Resin Reactor Specification 282
Table 94 Thermal Incinerator Operating Conditions for Resin Reactor Specification (Without Heat Exchange) 285

Page No.

Table 95 Estimated Capital Cost Data for Thermal Incinerators for Resin Reactors (Without Heat Exchange) 286
Table 96 Annual Operating Cost Data for Thermal Incinerators for Resin Reactors (Without Heat Exchange) 287
Table 97 Thermal Incinerator Process Description for Open Kettle Specification 288
Table 98 Thermal Incinerator Operating Conditions for Open Kettle Specification (Without Heat Exchange) 291
Table 99 Estimated Capital Cost Data for Thermal Incinerators for Open Kettles (Without Heat Exchange) 292
Table 100 Annual Operating Cost Data for Thermal Incinerators for Open Kettles (Without Heat Exchange) 293
Table 101 Catalytic Incinerator Process Description for Resin Reactor Specification 294
Table 102 Catalytic Incinerator Operating Conditions for Resin Reactor Specification (Without Heat Exchange) 297
Table 103 Estimated Capital Cost Data for Catalytic Incinerators for Resin Reactors (Without Heat Exchange) 298
Table 104 Annual Operating Cost Data for Catalytic Incinerators for Resin Reactors (Without Heat Exchange) 299
Table 105 Catalytic Incinerator Process Description for Open Kettle Specification 300
Table 106 Catalytic Incinerator Operating Conditions for Open Kettle Specification (Without Heat Exchange) 303
Table 107 Estimated Capital Cost Data for Catalytic Incinerators for Open Kettles (Without Heat Exchange) 304
Table 108 Annual Operating Cost Data for Catalytic Incinerators for Open Kettles (Without Heat Exchange) 305
Table 109 Thermal Incinerator Process Description for Resin Reactor Specification 310
Table 110 Thermal Incinerator Operating Conditions for Resin Reactor Specification (With Heat Exchange) 313
Table 111 Estimated Capital Cost Data for Thermal Incinerators for Resin Reactors (With Heat Exchange) 314
Table 112 Annual Operating Cost Data for Thermal Incinerators for Resin Reactors (With Heat Exchange) 315
Table 113 Catalytic Incinerator Process Description for Resin Reactor Specification 316
Table 114 Catalytic Incinerator Operating Conditions for Resin Reactor Specification (With Heat Exchange) 319
Table 115 Estimated Capital Cost Data for Catalytic Incinerators for Resin Reactors (With Heat Exchange) 320

Page No.

Table 116 Annual Operating Cost Data for Catalytic Incinerators for Resin Reactors (With Heat Exchange) 321
Table 117 Commercial Printing Market 333
Table 118 Ink Consumption in 1968 334
Table 119 Physical Properties of Commonly Used Solvents 337
Table 120 Percentage Breakdown of Solvent Consumed for Ink Dilution by Printing Process and Solvent Type (1968) 338
Table 121 Volume Breakdown of Solvent Consumed for Ink Dilution by Printing Process and Solvent Type (1968) 339
Table 122 Volume Breakdown of Total Solvent Consumption by Use and Solvent Type (1968) 340
Table 123 Percentage Breakdown of Total Solvent Consumption by Use and Solvent Type (1968) 341
Table 124 Analysis of Dryer Exhaust Gas from Web Offset Publication 344
Table 125 Thermal Incinerator Process Description for Web Offset Printing Specification 364
Table 126 Thermal Incinerator Operating Conditions for Web Offset Printing Specification (Without Heat Exchange) 367
Table 127 Estimated Capital Cost Data for Thermal Incinerators for Web Offset Printing (Without Heat Exchange) ... 368
Table 128 Annual Operating Cost Data for Thermal Incinerators for Web Offset Printing (Without Heat Exchange) ... 370
Table 129 Catalytic Incinerator Process Description for Web Offset Printing Specification 372
Table 130 Catalytic Incinerator Operating Conditions for Web Offset Printing Specification (Without Heat Exchange) 375
Table 131 Estimated Capital Cost Data for Catalytic Incinerators for Web Offset Printing (Without Heat Exchange) ... 376
Table 132 Annual Operating Cost Data for Catalytic Incinerators for Web Offset Printing (Without Heat Exchange) ... 378
Table 133 Thermal Incinerator Process Description for Metal Decorating Specification 380
Table 134 Thermal Incinerator Operating Conditions for Metal Decorating Specification (Without Heat Exchange) .. 383
Table 135 Estimated Capital Cost Data for Thermal Incinerators for Metal Decorating (Without Heat Exchange) 384
Table 136 Annual Operating Cost Data for Thermal Incinerators for Metal Decorating (Without Heat Exchange) 386
Table 137 Thermal Incinerator Process Description for Metal Decorating Specification 388

Page No.

Table 138 Thermal Incinerator Operating Conditions for Metal Decorating Specification (With Heat Exchange) 391
Table 139 Estimated Capital Cost Data for Thermal Incinerators for Metal Decorating (With Heat Exchange) 392
Table 140 Annual Operating Cost Data for Thermal Incinerators for Metal Decorating (With Heat Exchange) 394
Table 141 Catalytic Incinerator Process Description for Metal Decorating Specification 396
Table 142 Catalytic Incinerator Operating Conditions for Metal Decorating Specification (Without Heat Exchange) .. 399
Table 143 Estimated Capital Cost Data for Catalytic Incinerators for Metal Decorating (Without Heat Exchange) 400
Table 144 Annual Operating Cost Data for Catalytic Incinerators for Metal Decorating (Without Heat Exchange) 402
Table 145 Catalytic Incinerator Process Description for Metal Decorating Specification 404
Table 146 Catalytic Incinerator Operating Conditions for Metal Decorating Specification (With Heat Exchange) 407
Table 147 Estimated Capital Cost Data for Catayltic Incinerators for Metal Decorating (With Heat Exchange) 408
Table 148 Annual Operating Cost Data for Catalytic Incinerators for Metal Decorating (With Heat Exchange) 410
Table 149 Thermal Incinerator Process Description for Gravure Printing Specification 412
Table 150 Thermal Incinerator Operating Conditions for Gravure Printing Specification (With Heat Exchange) 415
Table 151 Estimated Capital Cost Data for Thermal Incinerators for Gravure Printing (With Heat Exchange) 416
Table 152 Annual Operating Cost Data for Thermal Incinerators for Gravure Printing (With Heat Exchange) 418
Table 153 Carbon Adsorption Process Description for Gravure Printing Specification 420
Table 154 Carbon Adsorption Operating Conditions for Gravure Printing Specification 421
Table 155 Estimated Capital Cost Data for Carbon Adsorption for Gravure Printing 422
Table 156 Annual Operating Cost Data for Carbon Adsorption for Gravure Printing 424
Table 157 Confidence Limits for Capital Cost of Carbon Adsorption for Gravure Printing 426

Page No.

Table 158 Typical Operating Conditions for Continuous Sulfonating of Linear Alkylate 436
Table 159 Typical Synthetic Detergent Formulas 446
Table 160 Wet Scrubber Process Description for Soap and Detergent Spray Drying Specification 452
Table 161 Wet Scrubber Operating Conditions for Soap and Detergent Spray Drying Specification 453
Table 162 Estimated Capital Cost Data for Wet Scrubbers for Soap and Detergent Spray Drying 454
Table 163 Annual Operating Cost Data for Wet Scrubbers for Soap and Detergent Spray Drying 455
Table 164 Confidence Limits for Capital Cost of Wet Scrubbers for Soap and Detergent Spray Drying (Medium Efficiency) 460
Table 165 Confidence Limits for Capital Cost of Wet Scrubbers for Soap and Detergent Spray Drying (High Efficiency) 462
Table 166 Fabric Filter Process Description for Soap and Detergent Spray Drying Specification 464
Table 167 Fabric Filter Operating Conditions for Soap and Detergent Spray Drying Specification 465
Table 168 Estimated Capital Cost Data for Fabric Filters for Soap and Detergent Spray Drying 466
Table 169 Annual Operating Cost Data for Fabric Filters for Soap and Detergent Spray Drying 468
Table 170 Confidence Limits for Collector Only Cost of Fabric Filters for Soap and Detergent Spray Drying 470
Table 171 Fabric Filter Process Description for Soap and Detergent Product Handling Specification 472
Table 172 Fabric Filter Operating Conditions for Soap and Detergent Product Handling Specification 473
Table 173 Estimated Capital Cost Data for Fabric Filters for Soap and Detergent Product Handling 474
Table 174 Annual Operating Cost Data for Fabric Filters for Soap and Detergent Product Handling 476
Table 175 Confidence Limits for Collector Only Cost of Fabric Filters for Soap and Detergent Product Handling ... 478
Table 176 Typical Analyses of High Calcium and Dolomitic Commercial Limestones 488
Table 177 Typical Composition of High Calcium and Dolomitic Quicklime 489

Page No.

Table 178 Typical Composition of Reburned Quicklime 494
Table 179 Typical Compositions of Exhaust Gases from Lime Sludge Kilns 497
Table 180 Breakdown of Fuel Usage in Lime Rock Kilns During 196 498
Table 181 Typical Gaseous Sulfur Concentration Ranges and Average for Lime Sludge Kiln Exhaust 499
Table 182 Typical Emissions of a Natural Gas-Fired Lime Rock Kiln 501
Table 183 Typical Emissions of a Pulverized Coal-Fired Lime Rock Kiln 502
Table 184 Various Compositions of Lime Sludge Kiln Dust 503
Table 185 Typical Exhaust Gas Production for Several Lime Sludge Kiln Sizes 504
Table 186 Theoretical Lime Rock Kiln Exhaust Rates for Different Fuels 505
Table 187 Typical Particle Size Distribution of Vertical Lime Rock Kiln Emissions 507
Table 188 Wet Scrubber Process Description for Vertical Lime Rock Kiln Specification 512
Table 189 Wet Scrubber Operating Conditions for Vertical Lime Rock Kiln Specification 513
Table 190 Estimated Capital Cost Data for Wet Scrubbers for Vertical Lime Rock Kilns 514
Table 191 Annual Operating Cost Data for Wet Scrubbers for Vertical Lime Rock Kilns 515
Table 192 Confidence Limits for Capital Cost of Wet Scrubbers for Vertical Lime Rock Kilns (High Efficiency) 520
Table 193 Fabric Filter Process Description for Vertical Lime Rock Kiln Specification 522
Table 194 Fabric Filter Operating Conditions for Vertical Lime Rock Kiln Specification 523
Table 195 Estimated Capital Cost Data for Fabric Filters for Vertical Lime Rock Kilns 524
Table 196 Annual Operating Cost Data for Fabric Filters for Vertical Lime Rock Kilns 526
Table 197 Electrostatic Precipitator Process Description for Vertical Lime Rock Kiln Specification 528
Table 198 Electrostatic Precipitator Operating Conditions for Vertical Lime Rock Kiln Specification 529
Table 199 Estimated Capital Cost Data for Electrostatic Precipitators for Vertical Lime Rock Kilns 530

Page No.

Table 200 Annual Operating Cost Data for Electrostatic Precipitators for Vertical Lime Rock Kilns 531
Table 201 Confidence Limits for Capital Cost of Electrostatic Precipitators for Vertical Lime Rock Kilns (High Efficiency) 536
Table 202 Wet Scrubber Process Description for Rotary Lime Sludge Kiln Specification 538
Table 203 Wet Scrubber Operating Conditions for Rotary Lime Sludge Kiln Specification 539
Table 204 Estimated Capital Cost Data for Wet Scrubbers for Rotary Lime Sludge Kilns 540
Table 205 Annual Operating Cost Data for Wet Scrubbers for Rotary Lime Sludge Kilns 541
Table 206 Confidence Limits for Capital Cost of Wet Scrubbers for Rotary Lime Sludge Kilns (High Efficiency) 546
Table 207 Electrostatic Precipitator Process Description for Rotary Lime Sludge Kiln Specification 548
Table 208 Electrostatic Precipitator Operating Conditions for Rotary Lime Sludge Kiln Specification 549
Table 209 Estimated Capital Cost Data for Electrostatic Precipitators for Rotary Lime Sludge Kilns 550
Table 210 Annual Operating Cost Data for Electrostatic Precipitators for Rotary Lime Sludge Kilns 551
Table 211 Confidence Limits for Capital Cost of Electrostatic Precipitators for Rotary Lime Sludge Kilns (High Efficiency) 556
Table 212 Average Annual Foundry Production in the United States 563
Table 213 Properties of Cast Irons 565
Table 214 Exhaust Gases Compositions from Four Different Cupolas 574
Table 215 Material Balance for Theoretical Cupola 576
Table 216 Particle Size Distributions of Emissions from Three Different Cupolas 579
Table 217 Abatement Equipment Types Used for Collecting Particulate Matter From Foundries 580
Table 218 Fabric Filter Process Description for Gray Iron Foundry Cupola Specification 582
Table 219 Fabric Filter Operating Conditions for Gray Iron Foundry Cupola Specification 583
Table 220 Fabric Filter Operating Conditions for Gray Iron Foundry Cupolas 585

Page No.

Table 221 Estimated Capital Cost Data for Fabric Filters for Gray Iron Foundry Cupolas 586

Table 222 Annual Operating Cost Data for Fabric Filters for Gray Iron Foundry Cupolas 588

Table 223 Confidence Limits for Capital Cost of Fabric Filters for Gray Iron Foundry Cupolas 590

Table 224 Wet Scrubber Process Description for Gray Iron Foundry Cupola Specification 592

Table 225 Wet Scrubber Operating Conditions for Gray Iron Foundry Cupola Specification 593

Table 226 Estimated Capital Cost Data for Wet Scrubbers for Gray Iron Foundry Cupolas 596

Table 227 Annual Operating Cost Data for Wet Scrubbers for Gray Iron Foundry Cupolas 597

Table 228 Confidence Limits for Capital Cost of Wet Scrubbers for Gray Iron Foundry Cupolas (High Efficiency) 602

LIST OF APPENDICES

Appendix I Complete Sample Specification
Appendix II Instructions for Submitting Cost Data
Appendix III Labor Cost Indices
Appendix IV Statistical Basis for Data Presentation
Appendix V List of Standard Abbreviations

I. INTRODUCTION

The Industrial Gas Cleaning Institute (IGCI) is the national association of manufacturers of gas cleaning equipment, used primarily for the abatement of industrial air pollution. Under this twelve-month contract, the IGCI is collecting and formalizing data on air pollution control in seven industrial areas selected by the EPA. These areas are:

1. Phosphate Industry — Phase I
2. Feed and Grain Industry
3. Paint and Varnish Industry
4. Graphic Arts Industry — Phase II
5. Soap and Detergent Industry
6. Lime Kilns
7. Gray Iron Foundries — Phase III

This report contains all of the technical information assembled for each of the seven process areas. The technical material consists of a narrative description of each of the process areas tabulated above, specifications for air pollution abatement equipment for each, and a summary of capital and operating costs for equipment, obtained from the IGCI member companies in response to the specifications. The following section summarizes all of the technical data assembled for this study.

II. TECHNICAL DATA

This section contains all of the data collected as part of this program. This includes information on process descriptions, air pollution control requirements, specifications and capital and operating costs for abatement equipment used in these industries. Narrative material was generated by the combined efforts of Air Resources, Inc. personnel acting as editors and coordinators for the program, and the most qualified personnel of the member companies active in the field. The cost data, however, is entirely the product of companies judged most qualified. In addition to IGCI member companies, several non-members participated by supplying cost information. These companies prepared cost estimates independently of one another. Air Resources, Inc. consolidated the data and edited it with regard to format only.

GENERAL DESCRIPTION

1. FORMAT

This study includes seven industrial areas that were divided into three groups, each of which was covered by a separate phase report. In this final report, the three phases are summarized.

There are seven sections each covering one of the industrial areas. For each area, the following format is used:

1. Description of the Process
    a. Manufacturing or Production Aspects
    b. Air Pollution Control Equipment
2. Specifications and Costs
    a. Electrostatic Precipitators
        (1) Specifications
        (2) Capital Costs
        (3) Operating Costs
    b. Wet Scrubbers
    c. Fabric Filters
    d. Other
3. Summary Comments

This material will not be presented in outline form, nor will each item necessarily be included for each process area.

2. SELECTION OF APPLICABLE EQUIPMENT TYPES

In previous studies conducted by the IGCI, particulate emissions were of more concern than emissions of gaseous contaminants. In this study, there are three process areas in which the gaseous emissions predominate, or are equally important. These are:

Phosphate Industry – HF and SiF_4

Paint and Varnish – hydrocarbons

Printing – organic inks and solvents

Also included are four areas where particulate emissions predominate:

Feed and Grain

Gray Iron Foundries

Lime Kilns

Soap and Detergent

Because of the wider range of potential air pollutants covered, abatement devices of almost every conceivable type will be involved in the program. These will include all of the gaseous control devices:

Wet Scrubbers

Gas Incinerators

Adsorption Units

and the particulate control devices:

Electrostatic Precipitators

Wet Scrubbers

Fabric Filters

In general, a given process is amenable to control by more than one type of equipment. The Engineering Standards Committee of the IGCI has been responsible for selecting the types which will be considered in this program. In many areas, the EPA is conducting simultaneous programs in which industry surveys, source tests, and other programs may furnish additional insight into the equipment types predominating in well-controlled installations. This information was incorporated into the judgments reached by the Engineering Standards Committee through a series of technical exchange meetings with the EPA.

For the seven process areas covered by this study, selections of equipment types to be studied were made during three Engineering Standards Committee meetings and four technical exchange meetings. Present at the technical exchange meetings were representatives of the following groups:

EPA	Economic Analysis Branch
EPA	Industrial Studies Branch
EPA	Performance Standards Branch
IGCI	Technical Director
IGCI	Project Director
IGCI	Engineering Standards Committee
ARI	Project Coordinator
ARI	Project Engineer

The end results of this selection process are presented on Table 1.

3. BASIS FOR PREPARING SPECIFICATIONS AND BID PRICES

The degree of reduction of emissions required in a given application will influence the cost significantly for wet scrubbers and electrostatic precipitators. Fabric filters, mechanical collectors and incinerators are, on the other hand, relatively insensitive to the efficiency level specified. In all cases, the cost is directly related to size or gas handling capacity required.

In order to make a meaningful comparison of capital and operating costs, it is necessary to specify the performance level, or degree of reduction of emissions required. In previous projects, two arbitrary levels of performance were specified:

1. An "intermediate level" which corresponded to the Los Angeles process weight requirements, and

2. A "high level" of performance which should show little or no subjective evidence of emissions; that is no visible particulate matter, and no detectable odor level.

TABLE 1

ABATEMENT EQUIPMENT TYPES SELECTED FOR THE SEVEN PROCESS AREAS

Process Area	Emission Source	Equipment Types
Phosphates	Wet Phosphoric Acid Reactors Superphosphoric Acid Process	Packed Cross Flow Scrubber Water Induced Venturi Scrubber Venturi Scrubber with Packed Mist Eliminator
	Diammonium Phosphate Process	Two Stage Cyclonic Scrubber Venturi – Cyclone Scrubber Packed Cross Flow Scrubber
	Granular Triplesuperphosphate Reactors	Venturi – Cyclone Scrubber
	GTSP Storage	Cyclonic Scrubber
Feed and Grain	Feed Mills	Fabric Filter
	Flour Mills	Fabric Filter
	Feed Flash Dryers	Wet Scrubber, Thermal Incinerator
Paint and Varnish	Open Kettles	Thermal and Catalytic Incinerator
	Closed Reactors	Thermal and Catalytic Incinerator
Graphic Arts	Web Offset	Thermal and Catalytic Incinerator
	Metal Decorating	Thermal and Catalytic Incinerator
	Gravure Printing	Thermal Incinerator, Adsorption
Soap and Detergent	Spray Dryers	Fabric Filter, Wet Scrubber
	Dry Product Handling	Fabric Filter
Lime Kilns	Vertical Rock Kilns	Wet Scrubber, Fabric Filter, Precipitator
	Rotary Sludge Kilns	Wet Scrubber, Precipitator
Gray Iron Foundries	Cupola	Fabric Filter, Wet Scrubber

These arbitrary levels served reasonably well with regard to the specification of performance of particulate collection equipment. However, they did not serve the intended purpose; i.e., the development of a relationship between system cost and efficiency or performance level when:

1. gaseous pollutants were involved
2. when the pollutants were particulate but the plant was large or
3. the pollutants were particulate but the particles were large in diameter.

In the latter two cases, the "intermediate level" of performance was likely to be equally or more restrictive than the "high level".

In the case of gaseous emissions, the process weight basis for specifications and the visible emission basis are clearly inappropriate. In these circumstances, an alternative basis for setting performance level may be drawn from regulations in force or proposed by state or local agencies, but unanimity with regard to appropriateness is far less likely for gaseous contaminants than was the case for particulates.

In phosphate fertilizer plants, the emission of gaseous fluorides in the form of hydrogen fluoride and silicon tetrafluoride is the most critical problem. These plants are mostly located in the State of Florida, which has adopted stringent limitations on total fluoride emissions to protect sensitive citrus crops from damage by fluorides.

It was agreed by discussion with EPA personnel that the regulations adopted by the State of Florida and currently applicable to new plants (to be enforced on existing plants in 1975) would be used as the basis for the lower efficiency level, and one-half of that absolute emission rate would be used for the higher efficiency case.

For the six process areas covered in Phases II and III, performance levels were chosen during the technical exchange meetings. For those areas where the emissions are particulate matter, performance levels were generally chosen as follows:

Medium Efficiency — The process weight table published in the Federal Register April 7, 1971

High Efficiency — A sufficiently low grain loading to achieve a clear stack.

The process weight table used is presented in Table 2.

For those areas where the emissions are gaseous, performance levels were selected to correspond to the maximum practical abatement efficiencies currently available to the industry under study. Table 3 shows the list of performance levels selected for all six industries.

Several simplifications were made in the preparation of the specifications which have some bearing on the results which are reported here. These should be kept in mind when using the prices, operating costs, etc. The form of the specification for equipment may have an influence over the price quoted. Overly-restrictive specifications may add 5 to 10% to the equipment price without a corresponding increase in value received by the purchaser. In each of the cases presented in this report, prices are based on a specification which covers most of the conditions of purchase in an equitable way. Instead of writing each specification independently, the participants agreed upon the general terms and conditions to be specified, and these conditions were made identical for each specification. The final specification in each case was made by inserting one page of descriptive material and one page of operating conditions pertaining to the specific application into the standard format. To avoid unnecessary repetition, a sample of the complete specification for one of the applications is included as Appendix I to this report. Only the pages pertinent to specific applications are contained in the body of the report.

Prices were requested in such a way as to indicate three bases:

1. Air pollution control *device*. This includes only the flange-to-flange precipitator, fabric filter, scrubber, etc.

2. Air pollution control *auxiliary equipment*. This includes major items such as fans, pumps, etc.

3. Complete *turnkey installation*. This includes the design, all labor and materials, equipment fabrication, erection and startup.

In order to maintain a consistent approach to quoting in each area, the specifications were written around the air pollution control *device*. The process description was, however, made general enough to allow the members to quote on the auxiliary equipment, such as fans, pumps, solid handling devices, etc., and to quote on an approximate installation cost. A complete set of instructions for quoting is given in Appendix II.

Labor costs are a variable from one location to another, and it was not possible to establish the complex pattern of variations in turnkey prices which occurs as a function of local variations in hourly rate, productivity and availability of construction tradesmen. In order to provide a consistent basis for the preparation of price quotations, the cost indices given in Appendix III were used. This was taken from "Building Construction Cost Data, 1970". This gives a construction cost index for 90 cities, using 100 to represent the national average. These figures are for the building trades, but they should be representative of field construction rates in general.

(text continued on p. 10)

TABLE 2

PROCESS WEIGHT TABLE FOR MEDIUM EFFICIENCY CASES*

Process Weight Rate (lb/hr)	Rate of Emission (lb/hr)
100	0.551
200	0.877
400	1.40
600	1.85
800	2.22
1,000	2.58
1,500	3.38
2,000	4.10
2,500	4.76
3,000	5.38
3,500	5.96
4,000	6.52
5,000	7.58
6,000	8.56
7,000	9.49
8,000	10.44
9,000	11.2
12,000	13.6
16,000	16.5
18,000	17.9
20,000	19.2
30,000	25.2
40,000	30.5
50,000	35.4
60,000 or more	40.0

---

*Federal Register April 7, 1971

TABLE 3

PERFORMANCE LEVELS SELECTED FOR PHASE II AND III PROCESS AREAS

Process Area	Emission Type	Performance Level Medium	 High
Feed and Grain			
Feed Mills	Particulate	(1)	0.01 gr/ACF
Flour Mills	Particulate	(1)	0.01 gr/ACF
Feed Flash Dryers	Odor	67% Removal	93% Removal
Paint and Varnish			
Open Kettles	Hydrocarbon Gas	(1)	1600° F Maximum Incinerator Temp.
Closed Reactors	Hydrocarbon Gas	(1)	1600° F Maximum Incinerator Temp.
Graphic Arts			
Web Offset	Hydrocarbon Gas	(1)	1600° F Maximum Incinerator Temp.
Metal Decorating	Hydrocarbon Gas	(1)	1600° F Maximum Incinerator Temp.
Gravure Printing	Hydrocarbon Gas	(1)	1600° F Maximum Incinerator Temp.
Soap and Detergent			
Spray Dryers	Particulate	PWT (2)	0.01 gr/ACF
Dry Prod. Handling	Particulate	(1)	0.01 gr/ACF
Lime Kilns			
Vertical Rock	Particulate	PWT (2)	0.01 gr/ACF
Rotary Sludge	Particulate	PWT (2)	0.01 gr/ACF
Gray Iron Foundries			
Cupolas	Particulate	PWT (2)	0.02 gr/ACF

(1) Only one efficiency level specified.

(2) PWT = Process Weight Table, See Table 2.

These figures do not take productivity differences into account and may understate the variations in cost from one city to another.

The participating companies were instructed to estimate the installation costs as though erection or installation of the system would be in Milwaukee, Wisconsin or another city relatively convenient to the participants' point of shipment with a labor rate near 100. Readers are cautioned to take local labor rates and *productivity* into account when making first estimates of air pollution control system installed costs based on the data in this report. Appendix III shows the tabulated hourly rates for various construction trades (based on national averages) which may be useful for this purpose.

Considerable emphasis was placed on estimation of operating costs. Manufacturers submitting costs for equipment were asked to estimate the operating costs in terms of utility requirements, maintenance and repair labor, and operating labor. These were requested in terms of the *quantity* required, rather than the cost. This is because the costs will be analyzed in terms of low, average and high utility and labor cost areas for the final report. For this report, only the average utility costs were used for preparing total annual cost figures.

4. PRESENTATION OF DATA

In general, the capital cost data is presented as a series of three graphs showing capital costs as a straight-line plot on log-log paper. One graph shows the capital cost for the air pollution abatement device, one shows the cost for the total equipment and one shows the cost for the complete turnkey system. Where it was possible to do so, an analysis of the confidence limits of the sample is presented — where three quotations were obtained as a "sample" of perhaps 20 possible suppliers who might have quoted. The confidence limits are shown as dotted lines. Appendix IV contains a description of the mathematical procedure involved.

In all specifications of pollution control equipment standard temperature is 70°F (530°R).

Operating costs are presented in graphic form in this report. Expanded tables showing operating costs for various utility cost levels are also included in some cases.

A total annual cost has been calculated for each process by combining an annual capital charge with a direct annual operating cost. The resulting figures are presented as a graph of total annual cost versus unit size where such information makes a meaningful pattern.

PROCESS DESCRIPTIONS, SPECIFICATIONS AND COSTS

1. PHOSPHATE ROCK PRODUCTS MANUFACTURE

Phosphorus is one of the elements essential to plant growth and animal life. For this reason, it is a major constituent of chemical fertilizers and animal feed supplements. The fertilizer-related uses of phosphorus have generated a large segment of industry involved with the mining and processing of natural phosphate rock deposits. This discussion includes this entire industry in a general way, and several specific manufacturing processes within the industry in detail.

Those sections of the phosphate rock processing industry chosen for detailed study are as follows:

a. Wet Process Phosphoric Acid manufacture (WPPA)

b. Superphosphoric acid manufacture (SPA)

c. Diammonium Phosphate (DAP)

d. Granular Triplesuperphosphate manufacture (GTSP)

e. GTSP Storage

Phosphate rock deposits exist in high concentration in only a few small geographic areas within the United States. Figure 1 is an outline map of the United States indicating those locations where phosphate rock deposits are found.(1) It is typical of the industry that the chemical processing plants which use phosphate rock as raw material and generate finished fertilizer products are located in the immediate vicinity of the mines and, therefore, within the geographic areas indicated in Figure 1.

The initial paragraphs in this section will be devoted to a generalized description of the over-all processing pattern within the phosphate fertilizer industry. Subsequent sections are related to the specifics of each process.

GENERAL DESCRIPTION

Phosphate rock is generally found in rich deposits of *fluorapatite,* with related minerals as impurities. Fluorapatite $Ca_{10}F_2(PO_4)_6$ is extremely stable and insoluble in water. For this reason, the phosphate content is not readily available for use by plants, nor is it leached from the deposits by percolation of water. In order to convert the rock to useful fertilizers, it is necessary to alter the chemical structure in such a way that the materials become at least moderately water soluble, or, more precisely, soluble in plant juices. Several chemical processes are used for converting the rock to soluble forms, three of which are among the specific subjects of this discussion. The processes in current use are:

(1) Phosphoric acid manufacture { wet process and electric furnace

(2) Normal Superphosphate

(3) Diammonium Phosphate

(4) Triplesuperphosphate

(5) Superphosphoric Acid

This discussion is concerned principally with those processes likely to be incorporated in a modern fertilizer production facility. These are generally arranged as shown in the block diagram in Figure 2. Ordinarily, ore benefication (upgrading) to produce a concentrate with 70 to 75% $Ca_3(PO_4)_2$ content,(3) or "70-75% *Bone Phosphate of Lime*". The scheme shown is tailored to the processing of Florida pebble phosphate rock, which comprises the world's largest source of phosphates.

Each of the processing steps involves some potential for the release of fluorine compounds into the air, and in some cases particulate solids as well. The description of the process sequence and basic chemistry has been oversimplified for this general description and the discussion of emission sources and control methods deferred to the sub-sections dealing with each specific process.

Phosphoric Acid Manufacture

The manufacture of phosphoric acid is involved with decomposition of the fluorapatite by sulfuric acid. This dissolution or digestion process results in the release of much of the fluorine as hydrofluoric acid or fluosilicic acid. Some of this is discharged from the process and becomes a potential for air

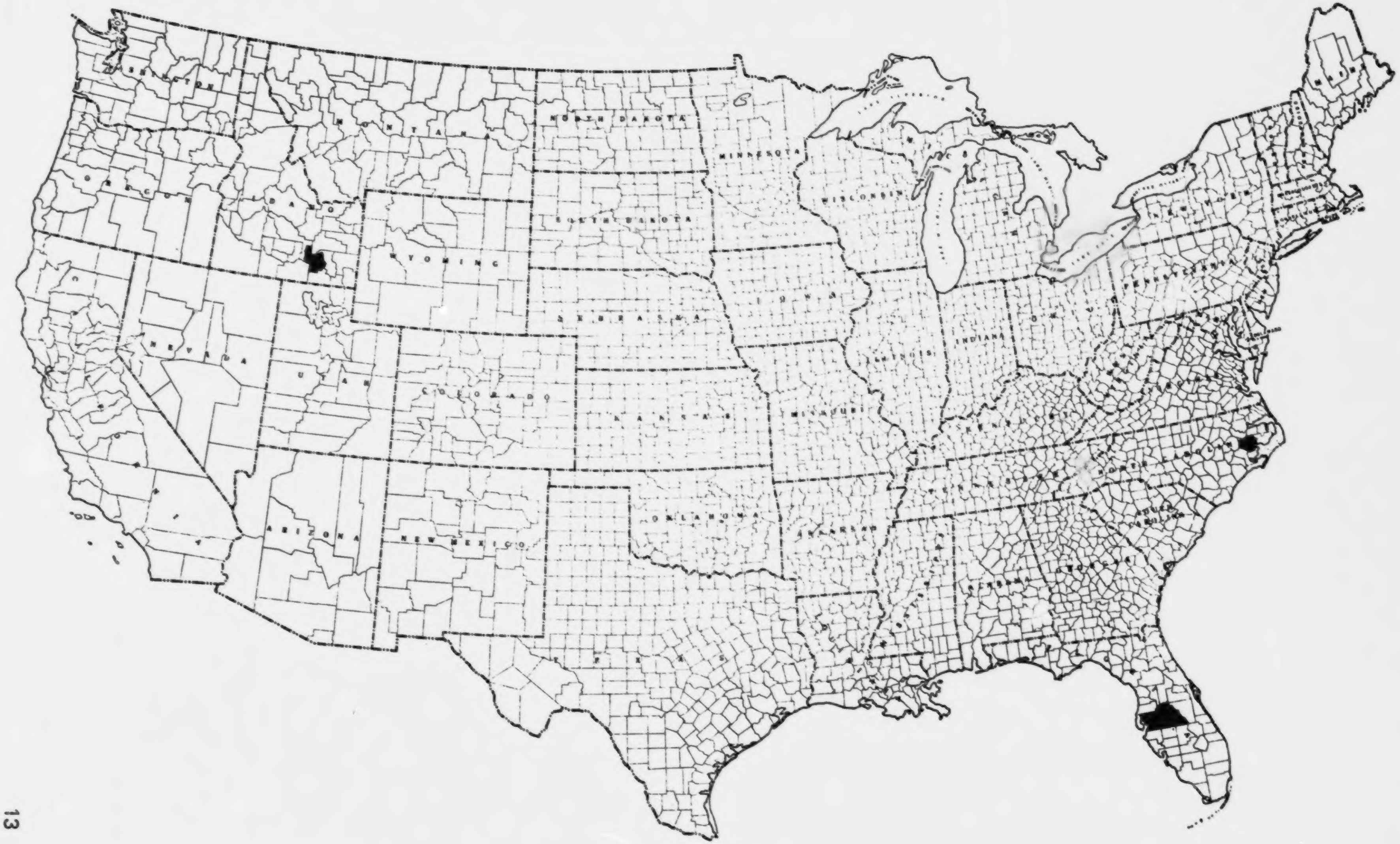

WASHINGTON
MONTANA
NORTH DAKOTA
MINNESOTA
OREGON
IDAHO
WISCONSIN
SOUTH DAKOTA
NEW YORK
MAINE
WYOMING
IOWA
PENNSYLVANIA
NEBRASKA
NEVADA
ILLINOIS
INDIANA
OHIO
UTAH
COLORADO
KANSAS
MISSOURI
KENTUCKY
TENNESSEE
NORTH CAROLINA
OKLAHOMA
ARKANSAS
SOUTH CAROLINA
ARIZONA
NEW MEXICO
ALABAMA
GEORGIA
TEXAS

FIGURE 1

OUTLINE MAP OF U.S. INDICATING
PHOSPHATE PRODUCING AREAS

pollution, some is retained in the acid product, and some is discarded with byproduct gypsum. The basic chemistry of the manufacture of phosphoric acid is represented by the following equation:

$$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \rightarrow 3\,[CaSO_4 \cdot 2H_2O] + 2H_3PO_4 \quad (1)$$

Phosphate Rock + Sulfuric Acid + Water →Gypsum + Phosphoric Acid

Although some dust is generated by processing of dry rock, release of gaseous fluorine-containing compounds constitutes the principal air pollution sources which must be treated in a phosphoric acid plant.

It is unusual to apply phosphoric acid directly as a fertilizer component, but it is more generally used in the manufacture of other phosphorus-containing compounds for direct application as fertilizers.

TSP Manufacture

Triplesuperphosphate is one of the common fertilizer forms marketed. It has a good balance of properties, including moderate water solubility, lack of either strong alkalinity or basicity, and ease of handling. Triplesuperphosphate is made by reacting phosphate rock with phosphoric acid according to the following chemical reaction:

$$Ca_3(PO_4)_2 + 4\,H_3PO_4 + 3H_2O \rightarrow 3\,[CaH_4\,(PO_4)_2 \cdot H_2O] \quad (2)$$

Phosphate Rock + Phosphoric Acid + Water →Triplesuperphosphate

In the acidulation process, part of the elemental fluorine contained in the phosphate rock is released and again presents a potential for air pollution by HF and by SiF_4. In addition, both the rock charge material and the finished products are granular solids which have significant dusting problems associated with the bulk handling steps in the process

Earlier processes for the digestion of phosphate rock involved the production of normal superphosphate by acidulation with sulfuric acid, and the production of "run of pile" triplesuperphosphate. Both of these processes are still widely used in existing fertilizer plants, but do not represent the best current technology. The normal superphosphate process utilizes sulfuric acid to produce the monocalcium phosphate rich material according to

$$Ca_3(PO_4)_2 + 2H_2SO_4 + 4H_2O \rightarrow CaH_4(PO_4)_2 + 2(CaSO_4{\cdot}2H_2O) \quad (3)$$

Phosphate Rock + Sulfuric Acid + Water →Monocalcium Phosphate + Gypsum

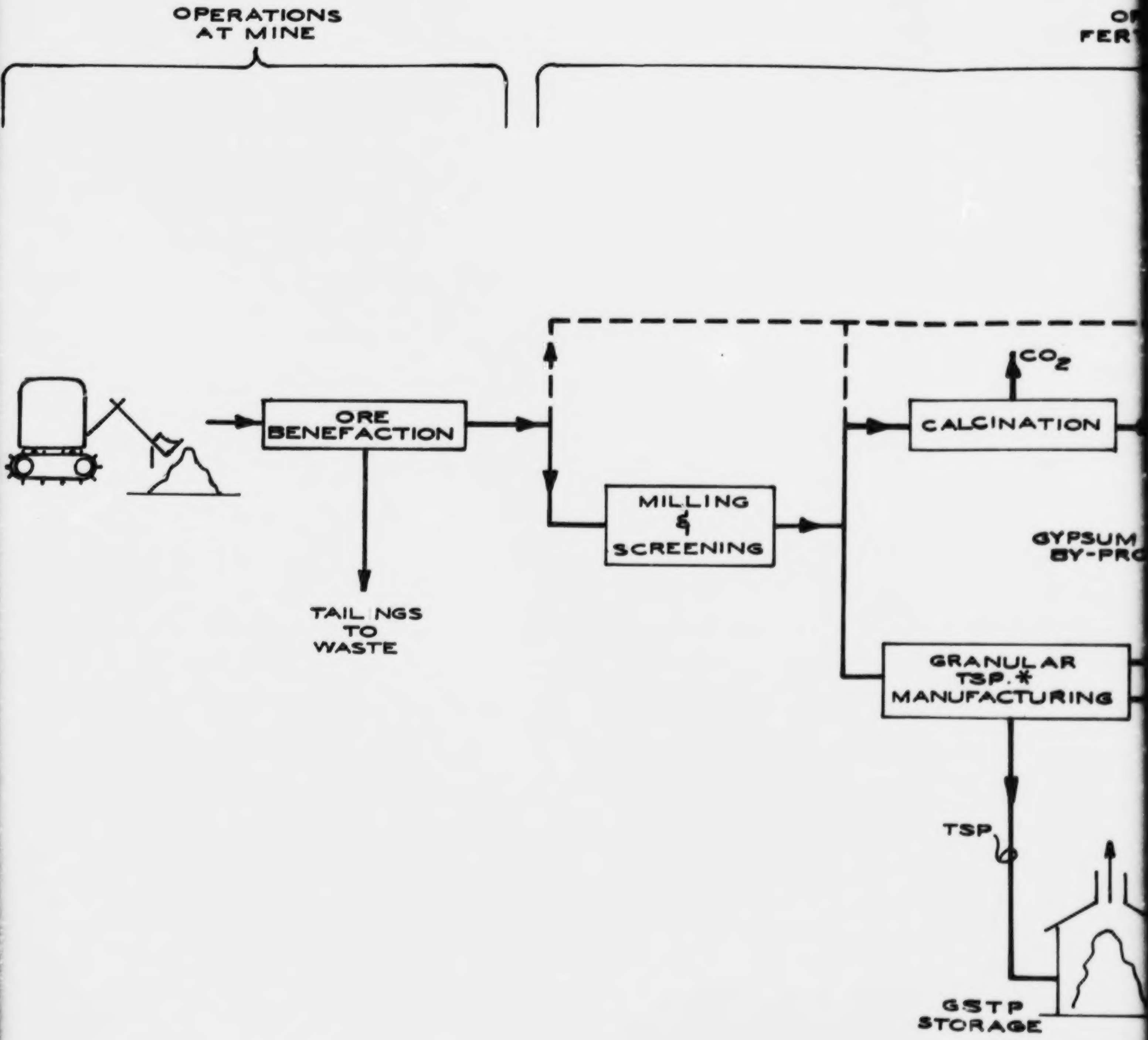
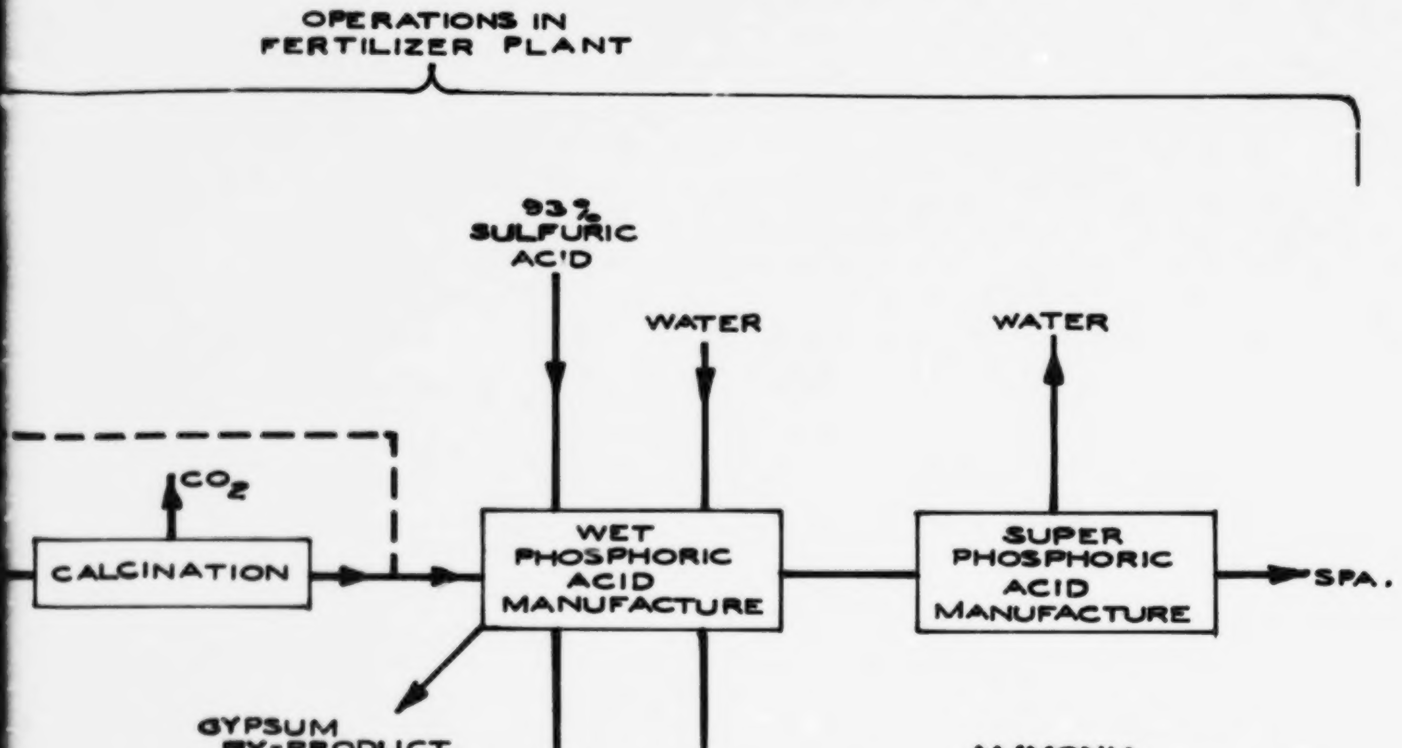

OPERATIONS AT MINE
OF
FERT
ORE BENEFACTION
TAILINGS TO WASTE
MILLING & SCREENING
CO_2
CALCINATION
GYPSUM
BY-PRO
GRANULAR TSP.* MANUFACTURING
TSP
GSTP STORAGE

FIGURE 2

BLOCK DIAGRAM OF PHOSPHATE
MANUFACTURING PROCES


OPERATIONS IN
FERTILIZER PLANT
93%
SULFURIC
ACID
WATER
WATER
CO$_2$
CALCINATION
WET
PHOSPHORIC
ACID
MANUFACTURE
SUPER
PHOSPHORIC
ACID
MANUFACTURE
SPA.
GYPSUM
BY-PRODUCT.
AMMONIA
GRANULAR
TSP.*
MANUFACTURING
GRANULAR
DAP.*
MANUFACTURING
TSP
GSTP
STORAGE
DAP
STORAGE

* USUALLY GTSP AND GDAP PRODUCTION ARE ALTERNATIVES WITHIN A GIVEN PLANT

FIGURE 2

AGRAM OF PHOSPHATE FERTILIZER
ANUFACTURING PROCESSES

This process has been supplanted by TSP manufacture because the TSP is not diluted by gypsum as is normal superphosphate.

The initial processes for manufacture of TSP utilized the "den" method, where solidification and curing of the product took place on a slow-moving belt. The chemistry of this process is substantially the same as that for the GTSP process, but the emission sources are significantly different. The physical properties of the product, called "run of pile" are less desirable than the granular product, and the latter is expected to be predominant in the future.

DAP Manufacture

Diammonium phosphate is a desirable component of fertilizers which require a high ratio of nitrogen to phosphorus. In this process, ammonia and phosphoric acid are used as raw materials, and produce diammonium phosphate as the principal product, with HF and SiF_4 as byproducts. The basic chemistry of the process is described by the following equation:

$$H_3PO_4 + 2NH_3 \rightarrow (NH_4)_2 HPO_4 \quad (4)$$

Phosphoric Acid + Ammonia → Diammonium Phosphate

NPK Manufacture

Fertilizers are usually blends of compounds containing available nitrogen, available phosphorus and available potassium. The proper application of fertilizers requires that these nutrients be supplied to the soil in the proportions required for balanced plant nutrition. Soil tests are run to establish the deficiency of each component in terms of tons/acre requirement for each element. Commercially available fertilizers are then applied on the basis of weight percent specifications for each nutrient in the particular fertilizer blend. The composition of the fertilizer is represented by the manufacturer in terms of pounds of each nutrient available per 100 pounds of total material. In order to make the proper application of elements to the soil in a particular area, it is usually necessary to apply several fertilizer blends. For this reason, it is necessary that fertilizer manufacturers make available a wide variety of blends of nitrogen, phosphorus and potash containing materials for the agricultural industry. Several of the products described in this study are high phosphate-content materials which have little or no nitrogen or potash content. Diammonium phosphate contains both nitrogen and phosphate, but most

Preceding page blank

compounded fertilizers are simply blends of phosphate materials with other chemicals which are rich in nitrogen or rich in potash. The plants in which this blending operation is carried out are called NPK plants, and carry out the chemical operations of mixing, blending, granulating and product handling associated with combining these elements into the commercial fertilizers in the proper proportions.

Storage and Handling

Because both raw materials and products of phosphate fertilizer plants are granular solids, there is a significant potential for air pollution emissions in the form of dusts at points where handling, transferring, milling, drying, storing and bagging of the solids are carried out.

Although these potential sources are important ones, this study has been restricted principally to those from which potentially serious gaseous fluoride emissions are likely. The only non-process source which falls into this category is the storage and curing buildings in which the granular TSP or DAP products are stored, cured and sometimes bagged. Within these buildings, the reactions involved in the final manufacturing steps are completed, and some emission of fluorine, moisture and heat occur. High ventilation rates through the storage buildings are required to maintain reasonable working conditions within the building. The ventilation air discharge from either DAP or TSP storage requires treatment to prevent significant emissions of fluorine-containing gases.

Summary of Applications

Prior to initiation of this program, representatives of the Industrial Studies Branch, the Performance Standards Branch, and the Economic Analysis Branch of the Environmental Protection Agency agreed upon the applications of equipment in the foregoing areas for which additional information on which technology and costs data should be assembled by the Industrial Gas Cleaning Institute.

Each of these applications is developed in some detail in the following sections.

PHOSPHORIC ACID MANUFACTURE

Phosphoric acid manufacture is carried out by the direct digestion of phosphate rock in sulfuric acid, or by the furnace method in which the rock is reduced to elemental phosphorus. Digestion in concentrated acid is the most widely used method for the production of phosphoric acid for use in fertilizers, while the electric furnace method is used to prepare high purity acid.(3) This discussion is limited to the sulfuric acid or wet process, frequently designated as the WPPA process.

Most of the wet process phosphoric plants in use are located in central Florida, where the product is used in the manufacture of phosphate-containing fertilizers. Several designs are in use, which differ in the detail but utilize nearly the same basic process flow scheme. This process is often described as the dihydrate process, because the byproduct gypsum formed is substantially all in the dihydrate form.

The chemical operations required for countercurrent washing, leaching and filtering of the rock involve steps which are difficult to carry out on a continuous basis, particularly because very corrosive acid materials are used. These problems have been solved by several ingenious designs.

PROCESS DESCRIPTION

Because the process flow scheme for a typical WPPA plant is relatively complex, a simplified flow diagram omitting many of the process steps is given in Figure 3. The discussion of basic process chemistry will refer to this flow drawing. Figure 4 is a more detailed diagram of the same process, showing an arrangement of process equipment which corresponds more closely with the operating plants.

In Figure 3, calcined, ground rock is introduced into a digester where it is slurried with a mixture of 24% phosphoric acid and 93% sulfuric acid. In the case of phosphoric acid, the acid strength refers to wt.% P_2O_5, while the sulfuric acid strength is given as wt.% H_2SO_4. Digestion of the rock is strongly exothermic, and requires heat removal to maintain a low and uniform digestion temperature, to keep the $CaSO_4$ byproduct from dissolving. The cooling is accomplished primarily by recycling a high slurry rate through a cooler chilled by vaporization of water and in some older plants by blowing cooling air through the digester. The reaction goes nearly to completion within the digestion chambers.

The slurry is agitated and circulated from compartment to compartment within the digester and a net stream of phosphoric acid-calcium sulfate slurry is withdrawn for filtering.

The filtering and washing operations are carried out on a series of filter surfaces shown symbolically as drum filters in Figure 3. Here the phosphoric acid-rich liquor entrapped within the solid $CaSO_4 \cdot 2H_2O$ crystal matrix is gradually replaced by washing with more and more dilute solutions, until, on the final filter, fresh make-up water is used for washing. The gypsum and entrained water are then sluiced to a gypsum settling pond.

The wash-water produced at each filtering-washing stage is recycled back into the process and contacted with solids of higher phosphoric acid content, so these streams build up in concentration and finally leave the first filter stage at about 30 wt.% phosphoric acid (throughout this discussion, "30 wt.% acid" refers to the percentage by weight of P_2O_5 rather than H_3PO_4. The "weight fraction of phosphoric acid" is, in this case, 41.5% H_3PO_4, which is equivalent to 30% P_2O_5).

A more detailed description of the process flow and equipment can be given using Figure 4 as a basis. Here, calcined phosphate rock is shown entering the process in three chambers of the digester along with 24 wt.% acid produced within the process and 93% sulfuric acid used as a raw material. These acids are mixed in a cooler which removes the heat of dilution of the H_2SO_4.

The slurry of rock and acid is pumped from chamber to chamber as the digestion reaction takes place with the concurrent solution of phosphate as H_3PO_4, silica as H_2SiF_6, precipitation of gypsum, release of HF from the fluorapatite, and volatilization of HF and SiF_4. The gaseous byproducts are swept away from the digester by the ventilation air stream.

The slurry of digested rock and acid progresses to a final digestion tank (shown as tank No. 9 in Figure 4) from which it is pumped to a circular flat pan filter where the countercurrent washing steps are carried out. The digestion tanks and filters are designed as complementary units in modern WPPA plants. Several versions of this design are in use, the most common of which is the Prayon design, built under license of S. A. Metalurgique de Prayon by Wellman-Powergas.(4) The filter is composed of flat, pie-shaped segments with slurry introduced above each section, and a vacuum drawn beneath the filter bed. The entire bed, consisting of about 8 segments, rotates beneath stationary feeding and washing connections, and after the solids have made one complete circuit, each filter segment is reversed top-for-bottom and the spent gypsum washed off for sluicing to the gypsum pond. This type of filter is known as a Bird-Prayon tilting pan filter.

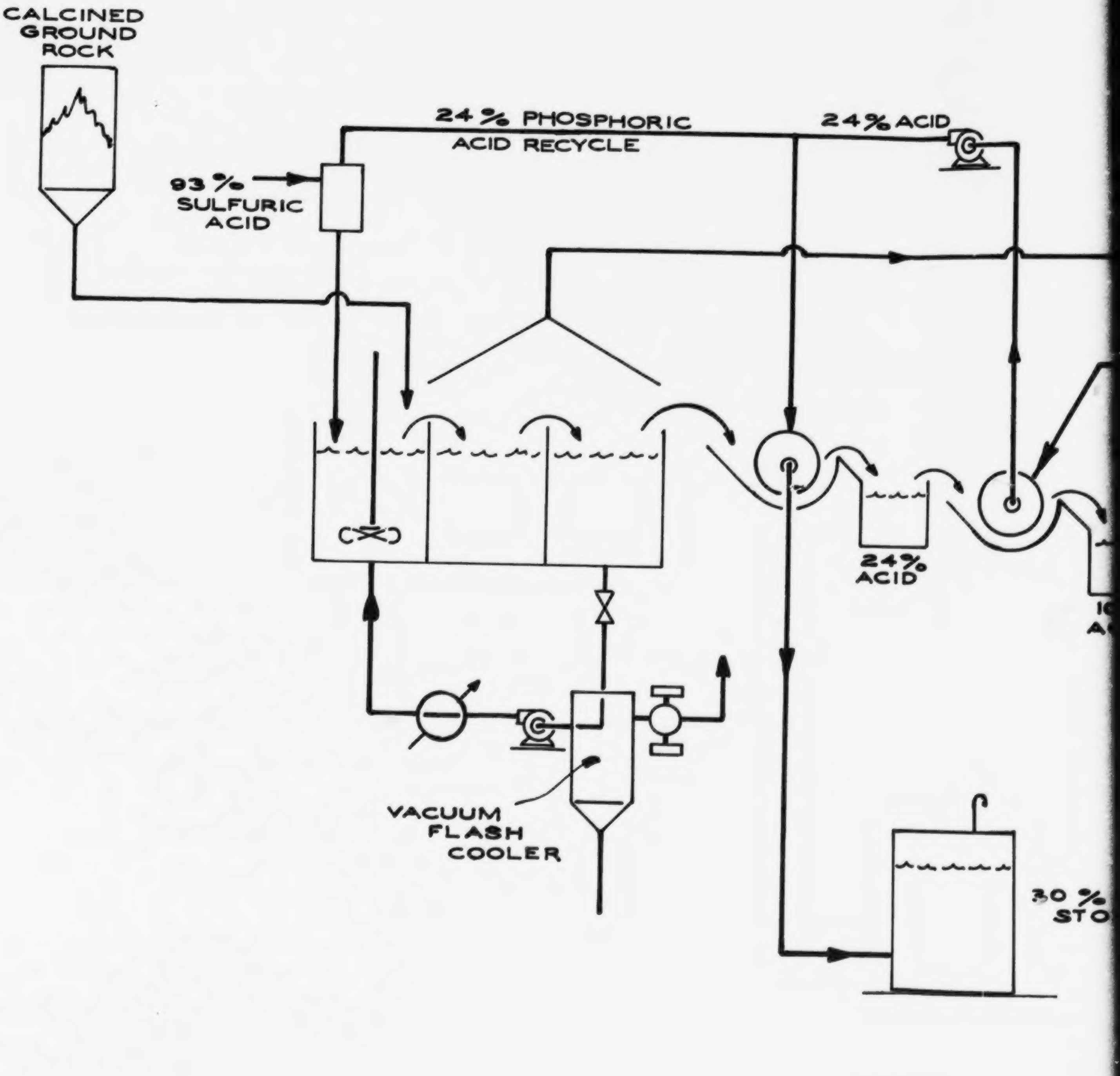
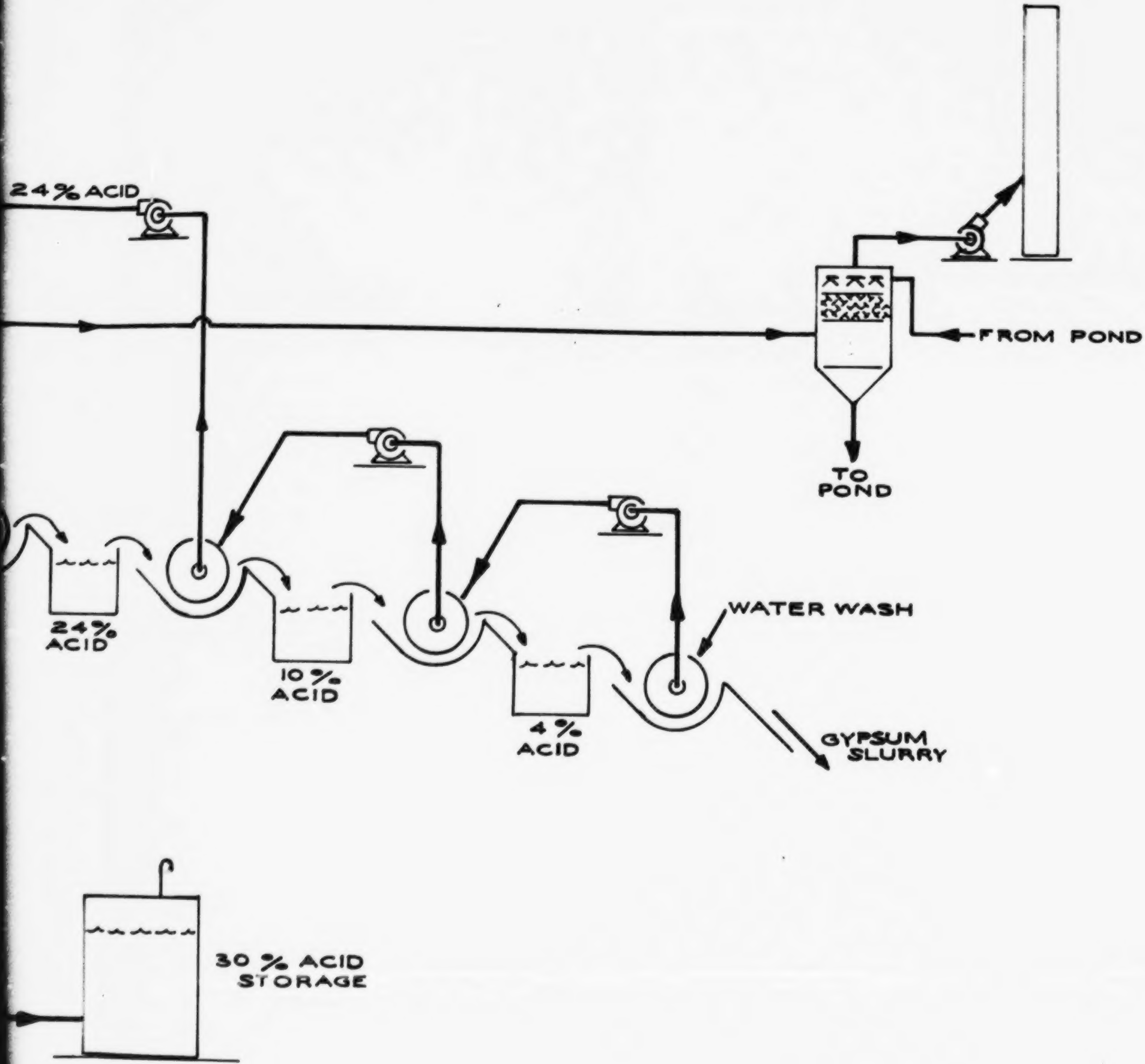

CALCINED GROUND ROCK
24% PHOSPHORIC ACID RECYCLE
24% ACID
93% SULFURIC ACID
24% ACID
VACUUM FLASH COOLER
30 %
STO

FIGURE 3

SIMPLIFIED FLOW SCHEME

FOR STRONG ACID PROCESS



24% ACID
FROM POND
TO POND
24% ACID
10% ACID
4% ACID
WATER WASH
GYPSUM SLURRY
30 % ACID STORAGE

FIGURE 3

IFIED FLOW SCHEME

TRONG ACID PROCESS

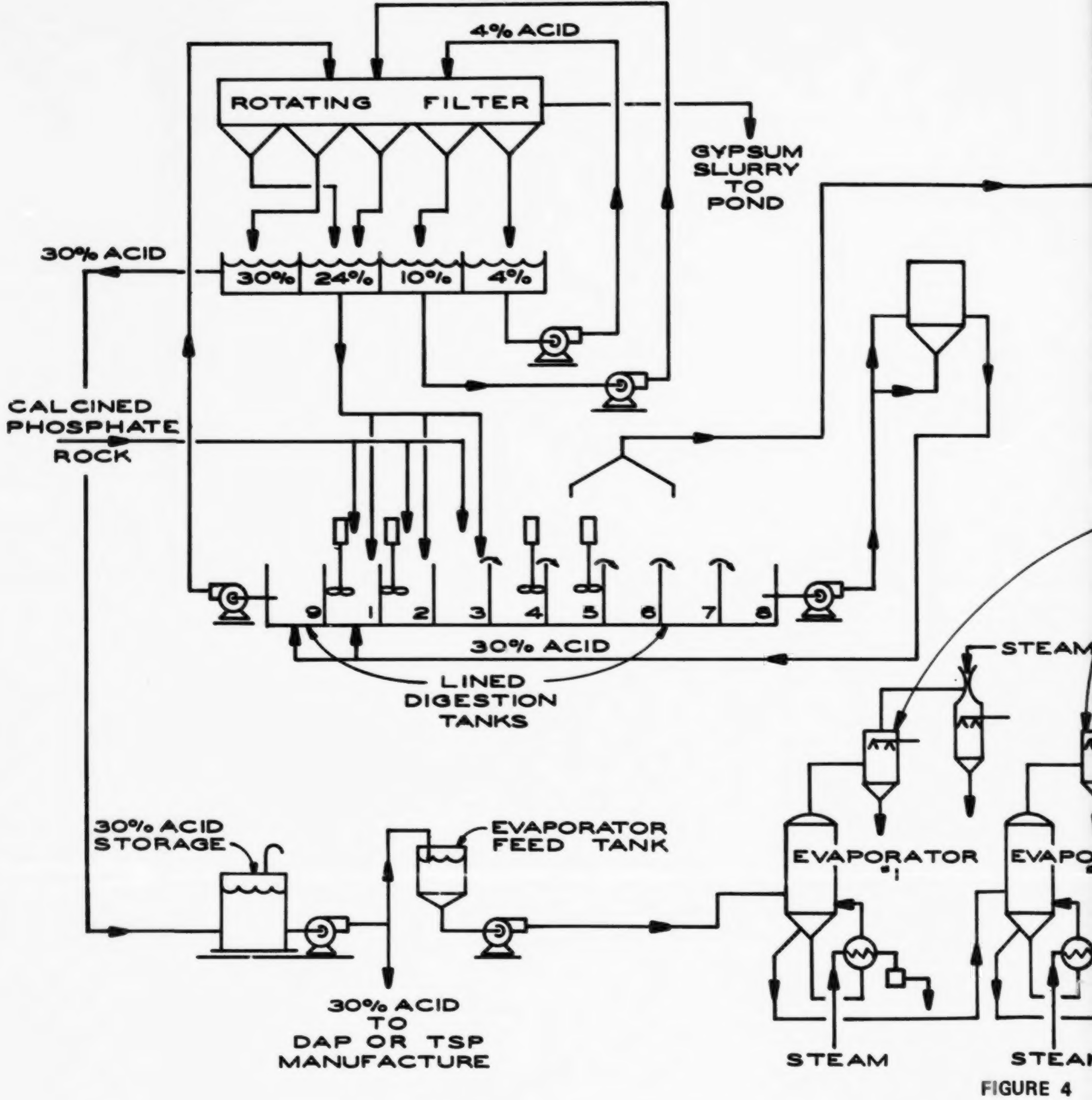

10% ACID
4% ACID
ROTATING FILTER
GYPSUM SLURRY TO POND
30% ACID
30%
24%
10%
4%
CALCINED PHOSPHATE ROCK
9
1
2
3
4
5
6
7
8
30% ACID
LINED DIGESTION TANKS
STEAM
30% ACID STORAGE
EVAPORATOR FEED TANK
EVAPORATOR #1
EVAPO
30% ACID TO DAP OR TSP MANUFACTURE
STEAM
STEA

FIGURE 4

SCHEMATIC FLOW DI

FOR WET PHOSPHORIC ACID
(GULF DESIGN PL

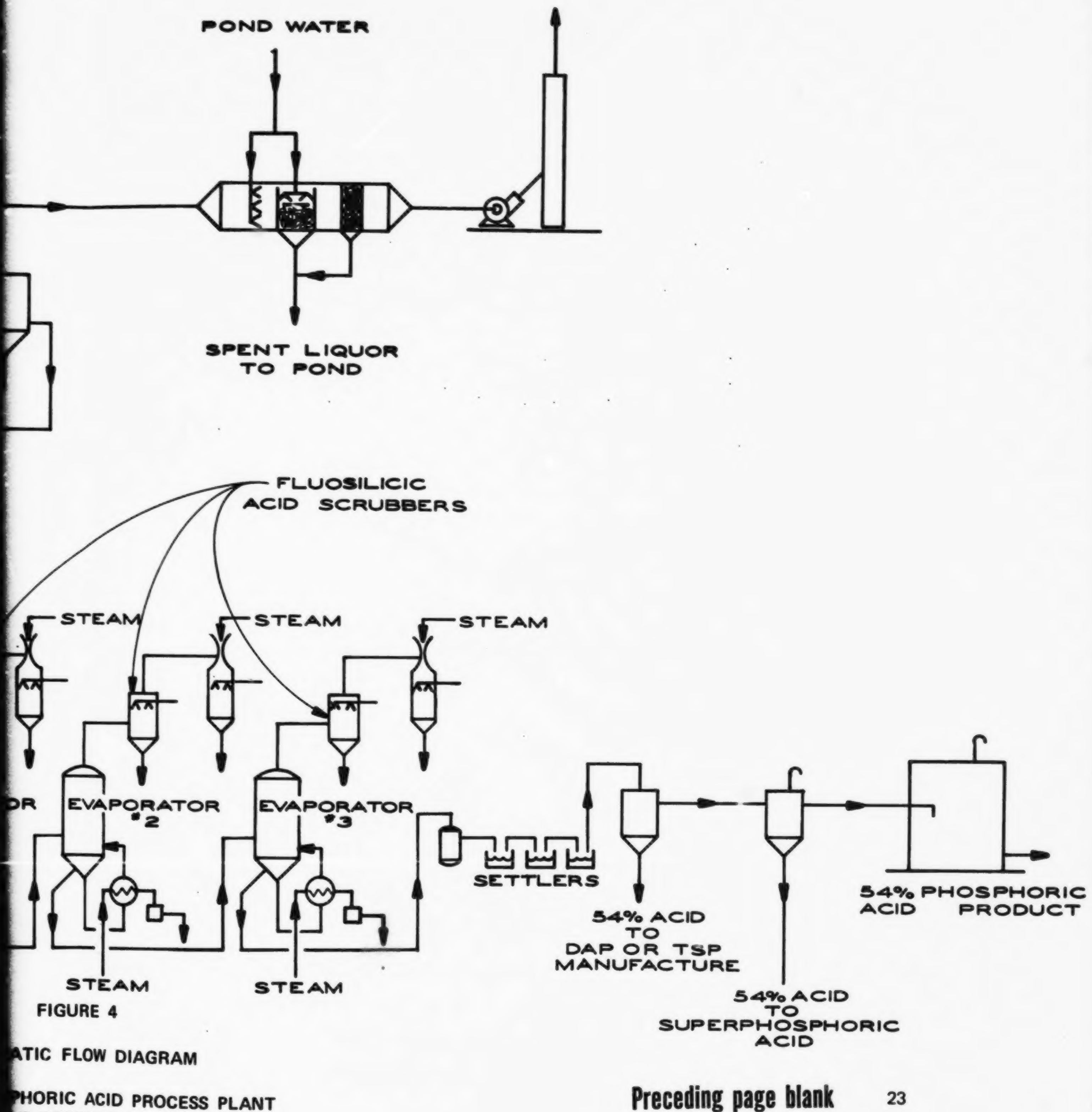

POND WATER
SPENT LIQUOR TO POND
FLUOSILICIC ACID SCRUBBERS
STEAM
STEAM
STEAM
OR
EVAPORATOR #2
EVAPORATOR #3
SETTLERS
STEAM
STEAM
54% ACID TO DAP OR TSP MANUFACTURE
54% ACID TO SUPERPHOSPHORIC ACID
54% PHOSPHORIC ACID PRODUCT

FIGURE 4

ATIC FLOW DIAGRAM

PHORIC ACID PROCESS PLANT
LF DESIGN PLANT)

Preceding page blank

Acid at 30% concentration from the rotating tilting-pan filter is discharged to a 30% acid storage tank from which part is removed for use in DAP or TSP manufacture and part is passed through a multiple effect evaporation system in which the concentration is raised to about 54% P_2O_5. The evaporators use high pressure steam to distill water overhead, and operate under vacuum to maintain low evaporation temperature. The vacuum is usually maintained by two-stage steam-jet ejectors with barometric condensers. In the concentration process, a substantial amount of fluorine is driven off as SiF_4. This is not ordinarily a significant air pollution source, in that most of the gas leaving the evaporators is steam rather than air and when the steam condenses in the barometric condensers the SiF_4 is combined with HF to form H_2SiF_6 and condenses along with it. However, if this material is allowed to escape to the gypsum pond, the potential for profitable recovery of the fluosilicic acid is lost and it places an additional chemical burden on the pond system.(5)

In Figure 4, each of the evaporators is shown equipped with a packed scrubber for the collection of H_2SiF_6, which can be produced at 20 wt.% concentration. One Florida processor reported the collection and sale of 5,000 T/year (on a 100 wt.% basis) of H_2SiF_6.(5)

Acid leaving the final evaporator is usually around 54 wt.% P_2O_5 (or 74.5 wt.% H_3PO_4).

The acid is passed through a series of clarifiers to remove as much as possible of the last traces of fines, and then discharged to storage. Product acid at 54% concentration may be taken for manufacture of DAP or TSP fertilizers prior to the final clarification stages. Acid for sale as product or for further concentration to superphosphoric acid may be subjected to several additional stages of clarification and may be centrifuged for final removal of fines.

CHEMISTRY OF PROCESS

The basic chemical reaction for the production of phosphoric acid by the dihydrate process as given in equation (1) is repeated here:

$$Ca_3(PO_4)_2 + 3H_2SO_4 + 6H_2O \rightarrow 3\,[CaSO_4 \cdot 2H_2O] + 2H_3PO_4 \quad (1)$$

Preceding page blank

This equation oversimplifies the reactions taking place in several respects. Most important from an air pollution standpoint is the presence of fluorine in the rock as fluorapatite. This may be considered to take part in a chain of reactions represented as follows:

$$3\,[Ca_3(PO_4)_2]\ \ CaF_2 \rightarrow 3Ca_3(PO_4)_2 + CaF_2 \tag{5}$$

Fluorapatite → Tricalcium Phosphate + Calcium Fluoride
or Fluorspar

$$CaF_2 + H_2SO_4 + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O + 2HF \tag{6}$$

Fluorspar + Sulfuric Acid + Water → Gypsum + Hydrofluoric Acid

If pure fluorapatite were the raw material for the digestion process, the principal gaseous byproduct would be HF, part of which would be retained in the acid and part would be liberated as gaseous HF. A major portion of the fluorine would probably be retained in the solid phase as CaF_2, which is extremely insoluble in water.

There are, however, some significant side reactions which have a great deal of influence on the disposition of the fluorine in the feed.

Principal among these are the reactions involving the solution of silica to form H_2SiF_6. A substantial amount of silica is present in the feed, either as free silica or as calcium, aluminum or iron silicates. Acidification of the rock brings about a series of reactions which may be represented as follows:

$$CaF_2 + H_2SO_4 \rightarrow CaSO_4 + 2HF \tag{7}$$

Calcium Fluoride + Sulfuric Acid → Calcium Sulfate + Hydrofluoric Acid

$$SiO_2 + 4HF \rightarrow 2H_2O + SiF_4 \tag{8}$$

Silica + Hydrofluoric Acid → Water + Silicon Tetrafluoride

$$SiF_4 + 2HF \rightarrow H_2SiF_6 \tag{9}$$

Silicon Tetrafluoride + Hydrofluoric Acid → Fluosilicic Acid

These reactions work to form HF, SIF_4 and H_2SiF_6. These three may be assumed to form an equilibrium system in the presence of water, from which the principal material escaping is SiF_4 in the normal range of digestion temperatures 130 - 170° F.(7)

FEED COMPOSITION

Several analyses of the typical Florida pebble phosphates are given in Table 4.(6) The analyses given do not report actual components, but rather convenient equivalents; that is, the sulfate-containing materials are not reported as such, but the total sulfur is reported as the equivalent SO_3. Such an analysis might be restated on the presumption that most of the phosphorus is present as fluorapatite, the CO_2 as carbonate, etc. Table 5 is a restatement of a typical analysis, first using the assumption that the principal impurities are present as common calcium compounds. A second restatement simply gives the analysis in terms of weight percent of each element.

The analyses shown in Table 5 are used subsequently as the raw material for hypothetical process plants defined for purposes of obtaining abatement equipment costs.

The reaction process is further complicated if $CaCO_3$ and organic materials present in most rock are allowed to enter the digester. Both give rise to serious foaming problems and require the addition of costly anti-foaming agents. Current practice usually avoids these problems by calcination of the rock in fluidized beds. Calcination effectively drives off CO_2 from the carbonates and oxidizes the organic material, without any significant loss of phosphorus or fluorine. Table 6 illustrates the effect of calcination on the feed composition.

The digestion reactions are aimed at producing the highest acid strength possible while producing crystals of gypsum byproduct which have good filtering characteristics. High temperature favors the production of high acid strength, but causes several operating problems, such as high rates of $CaSO_4 \cdot \frac{1}{2}H_2O$, or calcium sulfate hemihydrate, which is considerably more soluble than the dihydrate and tends to cause serious scaling problems as the acid cools. Other problems associated with high temperature operation of the digester are excessive corrosion and high solubility of impurities in the product acid. In most cases, the digester is controlled to about 180 - 185° F to produce a digester-acid with about 30% P_2O_5 content.(6)

Digester temperature is controlled either by blowing air into the slurry or by vacuum flash evaporation of a part of the slurry recycle. The use of air cooling is being supplanted by vacuum flash evaporation in the new plants.

TABLE 4

TYPICAL COMPOSITION AND PARTICLE SIZE OF COMMERCIAL GRADES OF FLORIDA PHOSPHATE ROCK[(a)]

	Specified BPL[(b)] Range				
	68/66	70/68	72/70	75/74	77/76
BPL (by analysis)	68.15	70.16	72.14	75.17	77.12
P_2O_5 equiv.	31.18	32.10	33.00	34.39	35.28
H_2O	1.3	1.0	1.0	1.0	1.0
Fe_2O_3	1.33	1.25	1.07	1.03	0.84
Al_2O_3	1.76	0.96	0.83	0.82	0.56
Organic	2.18	1.74	1.76	1.73	1.70
SiO_2	9.48	8.68	6.46	4.59	2.02
CO_2	3.48	3.05	2.87	2.65	2.98
F_2	3.60	3.67	3.62	3.78	3.89
CaO	45.05	46.12	48.10	50.14	51.53
SO_3	1.05	1.02	1.11	0.74	0.66

Screen Analysis

	% Larger or Smaller Than Indicated Mesh Size				
Specified Size[(c)]	+50	+70	+100	+200	−200
85% through 100 mesh	1.5	5.5	14.0	25.0	75.0
90% through 100 mesh	1.0	4.0	10.0	24.1	75.9

(a) Dry basis
(b) Bone Phosphate of Lime
(c) All grades

TABLE 5

ANALYSES OF PEBBLE PHOSPHATE ROCK

	Analysis Given Wt. %			Same Analysis Restated Wt. %		Same Analysis Restated Wt. %
BPL	(72.14)	$3Ca_3(PO_4)_2 \cdot CaF_2$		79.0	P	14.40
P_2O_5	33.0	$CaCO_3$		6.5	Ca	34.35
H_2O	1.0	CaF_2		1.3	Fe	0.75
Fe_2O_3	1.07	SO_3	(1)	1.1	Al	0.44
Al_2O_3	0.83	SiO_2	(1)	6.5	Si	3.01
Organic	1.76	Fe_2O_3	(1)	1.1	Co	2.40
SiO_2	6.46	Al_2O_3	(1)	0.8	O	40.17
CO_2	2.87	Oxygen	(1)	0.9	S	0.44
F_2	3.62	Organic	(2)	1.8	F	3.62
CaO	48.10	Water		1.0	H	0.24
SO_3	1.11			110.0		99.87
	99.82					

(1) Chemical combinations of these are uncertain.

(2) Assumed to be $(CH)_n$

TABLE 6

FEED TO WPPA PROCESS
BEFORE AND AFTER CALCINATION

	To Calciner		From Calciner	
Charge	T/D	lb/hr	T/D	lb/hr
(P_2O_5)	(520)	(43,330)	(520)	(43,300)
$3\ [Ca(PO_4)_2] \cdot CaF_2$	1,245	104,000	1,245	104,000
$CaCO_3$	103	8,600	–	–
CaF_2	20	1,710	20	1,710
CaO	–	–	58	4,800
Organics	28	2,340	–	–
Other	179	14,650	179	14,650
Total	1,575	131,300	1,502	125,160

Excessive sulfuric acid is necessary in the digester to precipitate the gypsum and assure near-complete solution of the phosphorus. Common practice with Florida rock is to maintain between 1.5 and 2.5% free H_2SO_4 by weight in the digester. This limits the loss of P_2O_5 to 0.2 - 0.7 wt.% of the total in the feed unreacted and discharged with the gypsum.(6) In addition to this loss, some additional material is lost by substitution of (HPO_4=) ions for (SO_4=) ions in the gypsum crystal lattice during washing,(2) and a loss of acid by entrainment in the filter cake. The overall efficiency of the process in converting P_2O_5 in the rock to phosphoric acid is about 95 to 97%.(6) Table 7 gives a typical composition of product acid.(6)

The distribution of impurities in the rock, and in particular of fluorine, is of concern in establishing the levels of pollutant concentration in the gas discharge.

Several references indicate typical concentration levels of impurities in the streams discharged from WPPA plants, or give percentage distributions of the total incoming fluorine. Table 8 summarizes these references. It is apparent that the total quantity of SiF_4 stripped out of the acid in the digester will vary with the digester temperature, the amount of cooling air used (which depends on the balance between cooling by air sparging or impingement and cooling by flash vaporization of slurry). The more cooling air used, the higher the fraction of fluorine stripped out and the lower the concentration.

Current practice minimizes the use of air for cooling the reactor, and as a consequence, the fluorine emission rate has been reduced below the values indicated in Table 8 by several-fold. The reactor cooling load is distributed principally between the sulfuric acid dilution cooler and the flash vaporization cooler:(15)

	MM BTU/ton P_2O_5
Sulfuric Acid Dilution Cooler	1.0
Flash Vaporization Cooler	1.3
Cooling Air	0.1
	2.4

TABLE 7

WET PROCESS PHOSPHORIC ACID

PRODUCT BREAKDOWN

	Wt. %	T/D	lb/hr
P_2O_5	54	500	41,690
F	1.5	14.3	1,190
SO_3	1.7	15.8	1,315
Al_2O_3	2.0	18.6	1,545
Fe_2O_3	0.8	7.4	620
Other	0.4	3.7	310
Water	39.6	367.2	30,610
Total	100.0	927.0	77,280

TABLE 8

DISTRIBUTION OF FLUORINE IN WPPA PLANT DISCHARGE STREAMS

Effluent	Percent of Total F By Weight		Concentration
	Given By Source	Used In Hypothetical Plant	
Rock Feed	100	100	3.6 wt.%[a]
Digester Vent Gas	5-10[d] 5[e]	1.5[f]	33-100[b] 200-500[c] 40-80[e] mg/SCF
Filter Vents	–	–	6[b] 10-30[c] mg/SCF
Sump Vents	–	–	0.5[b] 3-10[c] mg/SCF
Discharge With Gypsum	18-30[d] 30[e]	30	–
Collected by Evaporator Scrubber	35-50[d] 40[e]	40	–
Discharged With 54% Acid	25-35[d] 25[e]	28.5	2-5[c] wt.%

[a] To match example in Table II

[b] Reference (4)

[c] Reference (8)

[d] Reference (9)

[e] Reference (6)

[f] Chosen to match current practice.

The quantity of air drawn through the digester is established by the ventilation requirements rather than by the heat load.(15) The fluorine distribution in the second column of Table 8 was assumed for use in the remainder of this discussion.

A similar table could be prepared indicating the distribution of silica among the various discharges from the process; this distribution is considerably less critical from a pollution control standpoint, and it may simply be assumed that the gaseous fluorine discharges are all in the form of SiF_4 from the relatively low temperature digester, and H_2SiF_6 from the vacuum evaporators.

NATURE OF THE GASEOUS DISCHARGE

The most significant emission source in the typical WPPA process is the ventilating air from the digester. This is treated, in many cases, in combination with ventilating air streams from the filtration section of the plant and from the barometric condenser pumps. However, the digester emissions will be considered as a completely separate source in this study.

The digester vent gas consists of ambient air to which water vapor, particulate dust, and SiF_4 have been added as the air passes through the digester. Some old designs use air sparging through the slurry to accomplish the heat removal, while others employed jets of air blown down on the surface of the slurry. Recent designs do not rely upon the ventilating air to cool the reactor.

ABATEMENT REQUIREMENTS

The present Florida law, excerpted in Table 9, limits fluorine emissions from new WPPA plants to 0.02 lb/ton P_2O_5. In addition, old plants are required to conform to this restriction by July 1, 1975. In effect, this limits the emissions from a 500 T/D plant to 10 lb/day, or about 0.42 lb/hr. In each case the law uses the weight of P_2O_5 fed to the process.

By the most liberal interpretation, this would require the limitation of the digester effluent gas fluorine content to this value and require a removal efficiency of 99.82%. However, the overall emission specification *for the process* is 0.02 lb/T, and it is necessary to take into account the small amounts of fluorine in the gas streams vented over the filters and from the barometric condenser sumps in specifying the efficiency requirement for the digester.

Preceding page blank

TABLE 9

FLORIDA FLUORIDE EMISSION LAW

(c) Phosphate Processing — the emission limiting standards for phosphate processing are:

1. Fluorides (water soluble or gaseous-atomic weight 19) the following quantities expressed as pounds of fluoride per ton of phosphatic materials input to the system, expressed as tons of P_2O_5 for:

a. New plants or plant sections:

a 1. Wet process phosphoric acid production, and auxiliary equipment — 0.02 pounds of F per ton of P_2O_5.

a 2. Run of pile triple super phosphate mixing belt and den and auxiliary equipment — 0.05 pounds of F per ton of P_2O_5.

a 3. Run of pile triple super phosphate curing of storage process and auxiliary equipment — 0.12 pounds of F per ton of P_2O_5.

a 4. Granular triple super phosphate production and auxiliary equipment.

i. Granular triple super phosphate made by granulating run-of-pile triple super phosphate 0.06 pounds of F per ton of P_2O_5.

ii. Granular triple super phosphate made from phosphoric acid and phorphate rock slurry — 0.15 pounds of F per ton of P_2O_5.

a 5. Granular triple super phosphate storage and auxiliary equipment — 0.05 pounds of F per ton of P_2O_5.

a 6. Di ammonium phosphate production and auxiliary equipment — 0.06 pounds of F per ton of P_2O_5.

a 7. Calcining or other thermal phosphate rock processing and auxiliary equipment excepting phosphate rock drying and defluorinating — 0.05 pounds of F per ton of P_2O_5.

a 8. Defluorinating phosphate rock by thermal processing and auxiliary equipment — 0.37 pounds of F per ton of P_2O_5.

a 9. All plants, plant sections or unit operations and auxiliary equipment not listed in a.1 to a.8 will comply with best technology pursuant to Section 2.03(1) of this rule.

b. Existing plants or plant sections. Emissions shall comply with above section, 17-2.04(6)(c) 1.a., for existing plants as expeditiously as possible but not later than July 1, 1975 or

b 1. Where a plant complex exists with an operating wet process phosphoric acid section (including any items 17-2.04(6) 1., a., a.1. through a.6. above) and other plant sections processing or handling phosphoric acid or products or phosphoric acid processing, the total emission of the entire complex may not exceed 0.4 pounds of F per ton of P_2O_5 input to the wet process phosphoric acid section.

b 2. For the individual plant sections included in 17-2.04(6)(c), 1., a., 1.1. through 6.6 above but not included as a part as defined in 17-2.04(6)(c) 1., b., b.1 above, if it can be shown by comprehensive engineering study and report to the Department that the existing plant sections are not suitable for the application of existing technology, which may include major rebuilding or repairs and scrubber installations, the emission limiting standard to apply will be the lowest obtained by any similar plant section existing and operating.

Table 10 contains estimated gas flows, concentrations and loadings for all three gas streams, and illustrates the relationship of the digester vents to the other potential emission sources in the WPPA plant.

In addition to the gaseous fluorine in the digester vent, there is likely to be a little (0.05 gr/scf) rock dust generated by the mechanical handling of the calcined rock. This dust is quite likely to be removed incidentally to the removal of fluorine, however, a nominal specification for particulate removal efficiency is reasonable in that the rock contains some fluorine and will contribute to the overall discharge if allowed to escape from the process. A discharge of 0.005 gr/ACF of material with 3.5 wt.% F from a 500 T/D process would add

$$\frac{0.005}{7000} \times 25{,}000 \times 60 \times 0.035 = 0.037 \text{ lb/hr}$$

or nearly 10% of the allowed fluorine emission.

POLLUTION CONTROL EQUIPMENT

Of the types of pollution control equipment available, only those involving absorption (wet scrubbers) or adsorption (solid reagent or adsorption systems) are capable of removing gaseous fluorine compounds from ventilating air. Wet scrubbers have been used almost exclusively for this service although solid packed beds of limestone or alumina have been proposed for the removal of fluorine by adsorption.(11)

In this discussion only scrubbers have been considered as a part of the demonstrated technology.

Wet scrubbing combines the ability to remove matter from gas streams by impaction of the particulates on the surface of liquid droplets with the ability to absorb gaseous constituents into the liquid phase. Both of these functions are limited by the characteristics of the scrubbing liquor, the properties of the materials to be removed and sometimes by characteristics of the two in combination. In this case, the chemistry of the reactions between gypsum pond water and the fluorine-containing gases discharged from the WPPA plant reactor produce a gelatinous precipitate which tends to plug the packing in scrubbers and limits the types of scrubbing equipment applicable.

The basic chemistry of the compounds of fluorine, silicon and water must

TABLE 10

ESTIMATED ABATEMENT REQUIREMENTS

UNDER FLORIDA LAW

(for 500 T/D P_2O_5 WPPA plant)

	Digester	Filter	Sumps	Total
Gas Flow, SCFM	25,000	15,000	25,000	65,000
Fluorine Content				
mg/SCF	25	5.5	0.3	–
ppm	1,050	231	13	
lb/hr	80	11	1	99
Required Discharge				
mg/SCF	0.05*	0.05*	0.05*	0.05*
ppm	2.15	2.15	2.15	2.15
lb/hr	0.16	0.10	0.16	0.42
Efficiency Required, %	99.80	99.1	85	99.58

*Assumed equilibrium limited in scrubbing system.

be considered in order to characterize the application. In the reactor the fluorine contained in the fluorapatite or fluorspar goes into solution according to one or more of the following reactions:

$$CaF_{2(s)} + H_2SO_{4(1)} \rightarrow CaSO_{4(s)} + 2HF_{(1)} \qquad (10)$$

or

$$2HF_{(1)} + SiF_{4(1)} \leftarrow\rightarrow H_2SiF_{6(1)} \qquad (11)$$

A liquid vapor equilibrium is set up between the reactants in equation 11 above, which may be regarded as the reverse of equation 12.

$$H_2SiF_{6(1)} \rightarrow 2HF \uparrow + SiF_4 \uparrow \qquad (12)$$

High temperature tends to drive the reaction to the right, increasing the vapor pressure of both HF and SiF_4, and tending to increase the relative significance of silicon tetrafluoride as the fluorine-containing species. These vapor pressures set a lower limit of concentration in the gas phase leaving the scrubber. Although vapor pressure data for both HF and SiF_4 is available for the pure compounds, relatively little data is available in the literature on the vapor pressures of the two in water solutions. One Russian paper(12) does quote vapor pressures of both HF and SiF_4 and these are used throughout the remainder of this discussion.

Table 11 gives a conversion of the vapor pressure data to equilibrium ppm of SiF_4 and HF at various concentrations of H_2SiF_6 in the liquid phase. The last column in Table 9 gives the concentration of total fluorine in ppm F at 50°C.

Table 12 contains similar conversions for temperatures of 60 and 70°C. Table 11 data are plotted in Figure 5, which shows the distribution between fluorine vapor in equilibrium with acidic water at various temperatures. Figure 6 plots the total fluorine in the vapor with temperature as a parameter. Figure 7 is a cross plot of the data in Figure 6.

In addition to the reactions given, hydrolysis of SiF_4 occurs when the concentration of this component is higher than the equilibrium values, according to

$$3SiF_4 + 4H_2O \rightarrow Si(OH)_4 + 2H_2SiF_6 \qquad (13)$$

TABLE 11

CALCULATED CONCENTRATIONS OF SiF_4 AND HF AT 50°C

wt% H_2SiF_6	mmHg		ppm HF	ppm SiF_4	SiF_4 as ppm F	Total ppm F
	HF	SiF_4				
0.105	.0015	.0001	2.0	.1	0.4	2.4
0.55	.0018	.0002	2.4	.26	1.44	3.8
1.00	.0023	.0002	3.0	.26	1.44	4.4
2.64	.0030	.0003	4.0	.4	1.58	5.6
5.05*	.0050	.0004	6.6	0.5	2.10	8.7
5.05*	.0045	.0003	5.9	0.4	1.58	7.6
7.47*	.0045	.0011	5.9	1.4	5.80	11.7
7.47*	.0047	.0013	6.2	1.7	6.85	13.1
9.55	.0055	.0012	7.2	1.6	6.31	13.5
11.715	.0065	.0020	8.5	2.6	10.55	19.1

*Two experimental values were obtained at these concentrations.

TABLE 12

CALCULATED CONCENTRATIONS OF FLUORINE
IN VAPOR PHASE
AT 60°C AND 70°C

	60°C ppm				70°C ppm		
Wt.% H_2SiF_6	HF	4 x SiF_4	Total	Wt.% H_2SiF_6	HF	SiF_4 as ppm F	Total F
.105	2.24	1.58	3.82				
.55	3.30	1.05	4.35	0.55	6.58	2.64	7.22
.55	2.77	1.58	4.35	1.09	9.50	4.21	13.71
1.00	5.00	2.10	7.10	2.61	9.60	5.80	15.40
2.61	7.65	1.58	9.23	5.05	14.60	6.31	20.91
2.61	6.71	3.70	10.41	5.05	14.10	6.31	20.41
5.05	10.9	3.16	14.06	7.47	15.80	17.90	33.70
5.05	9.75	2.64	12.39	9.55	18.30	24.7	43.0
5.05	11.45	4.75	16.20	11.715	31.5	54.2	85.7
7.47	7.76	5.80	13.56	14.48	52.0	156.5	208.5
7.47	11.45	13.70	25.15				
9.55	10.8	14.75	25.55				
11.715	13.0	21.6	34.6				
14.48	28.0	55.5	83.5				

Preceding page blank

FIGURE 5

CONTRIBUTION OF HF AND SiF_4 TO VAPOR PHASE FLUORINE CONTENT AT 50°C

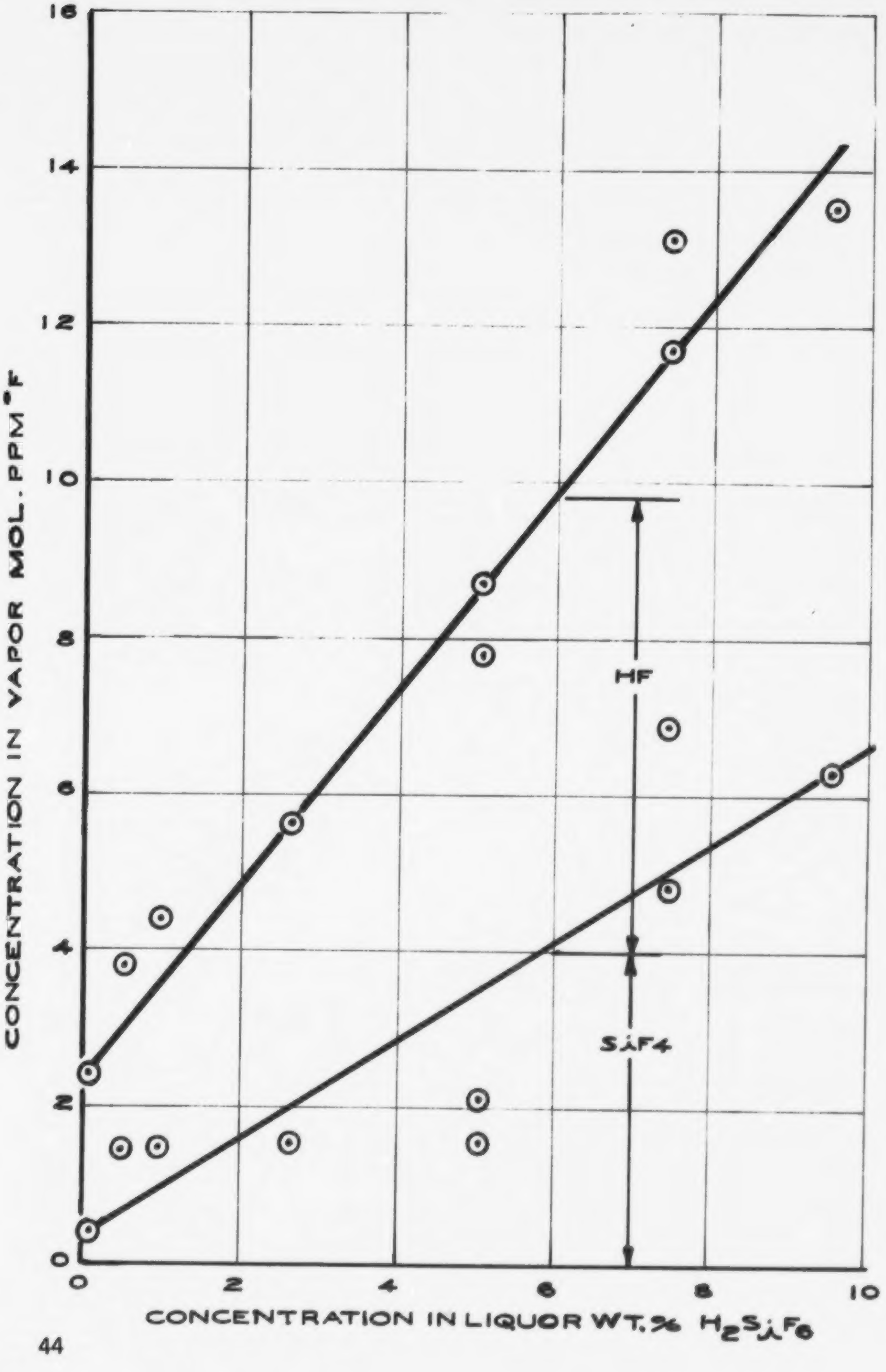


FIGURE 6

CONCENTRATIONS OF FLUORINE IN VAPOR OVER H_2SiF_6 SOLUTIONS

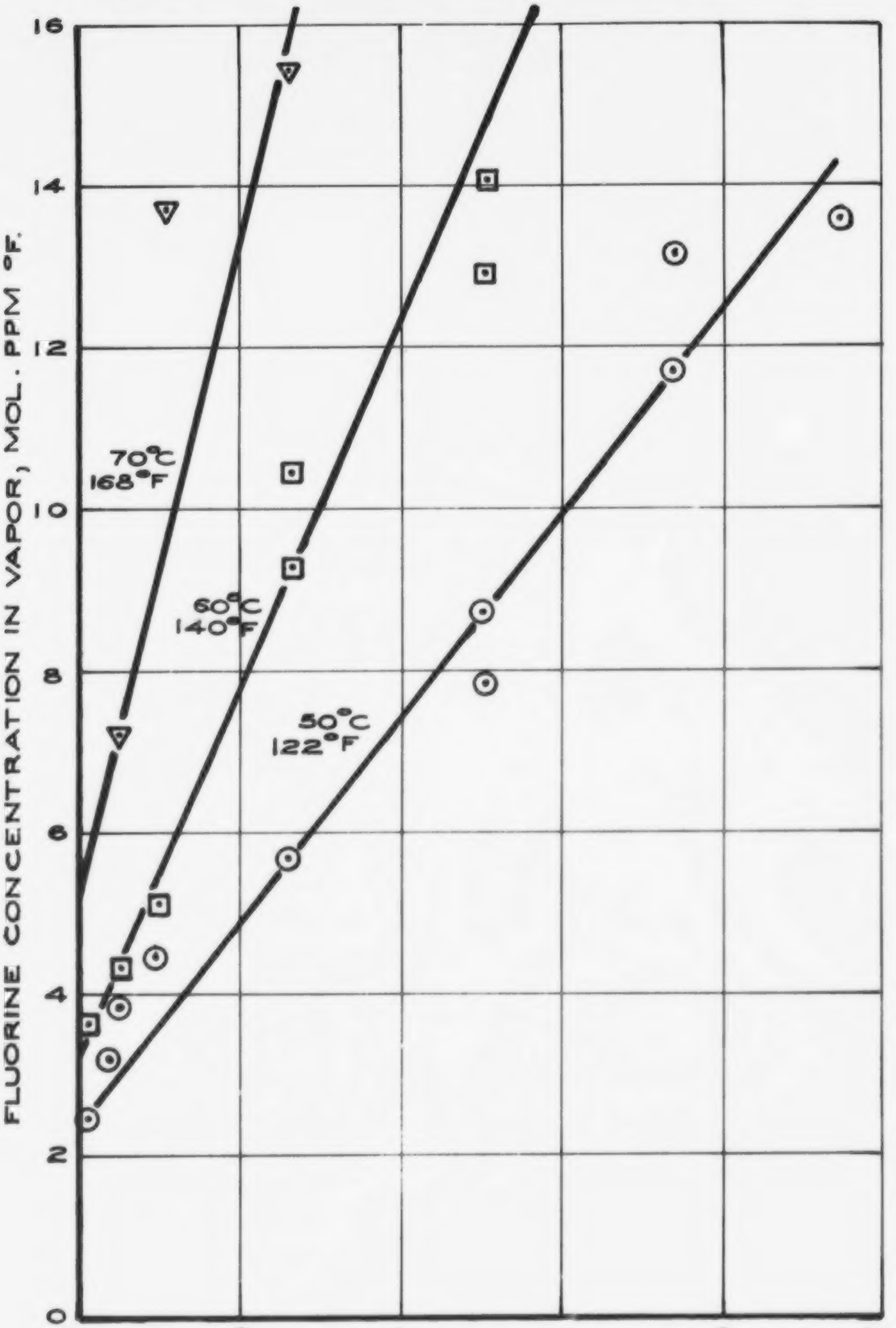
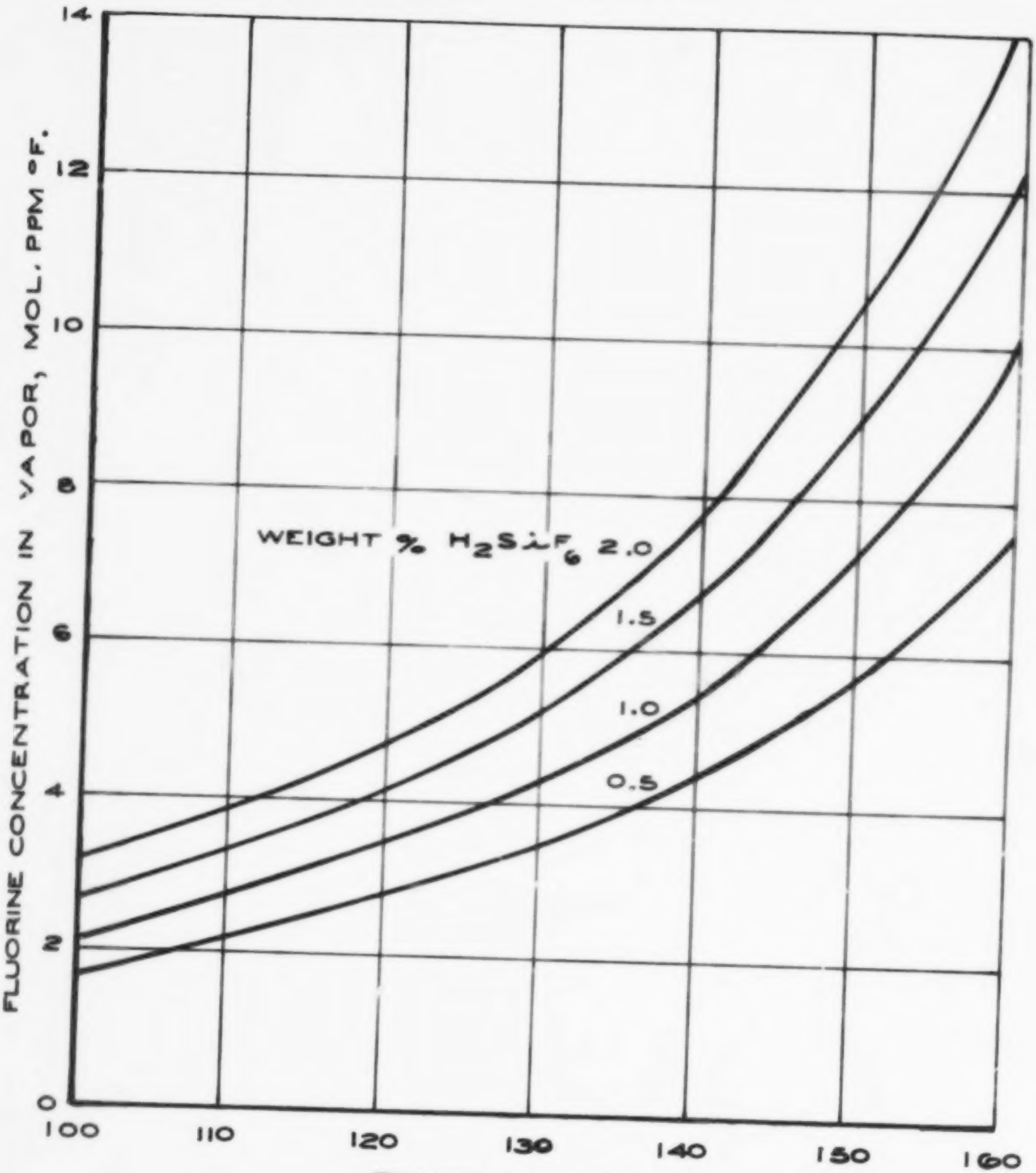


FIGURE 7

CONCENTRATION OF FLUORINE IN VAPOR OVER H_2SiF_6 SOLUTIONS VS. TEMPERATURE



This reaction tends to occur as the temperature of a gas stream is reduced in the presence of water, and leads to the formation of gelatinous deposits of polymeric silica which tend to plug scrubber packings. This problem limits the use of conventional packed countercurrent absorbers in this service, as well as other contacting devices which have small gas passages which might plug up. The remaining types of scrubbers which are likely to perform well in this service are

(1) Spray towers

(2) Wet cyclonic scrubbers

(3) Venturi scrubbers

(4) Concurrent packed absorbers

(5) Cross-flow packed absorbers

Spray towers are not capable of the high efficiencies required for compliance with present regulations. They may, however, be useful as pre-contactors to cool the gas stream and remove fluorine at relatively high concentration levels. They have relatively little pressure drop and can be used to bring large volumes of pond water into contact with the digester gas to reduce the temperature and improve the absorption equilibrium.

Wet cyclones are also limited in efficiency, but may be used as precoolers. They have a higher pressure drop requirement at high liquid flows than do the spray chambers.

Venturi scrubbers can bring about effective contact and gas absorption when sufficient energy is imparted to the gas to atomize the scrubbing liquor and create very small droplets. The contact time in a Venturi is very short, and it has been found that the power requirements at a given level of fluorine absorption are high as compared with packed scrubbers.(12)

Countercurrent scrubbers have an inherent advantage over concurrent or cross-flow scrubbers for gas absorption applications where the concentration of contaminant leaving the scrubber approaches equilibrium with the scrubbing liquor. This advantage is most clearly explained by reference to Figure 8.

Here the concentration of contaminant in the gas phase is plotted as a function of position in the scrubber. Y_1 represents the inlet concentration and

y_2 the outlet concentration. In a counter-current scrubber, Figure 8a, the liquor contains some of the contaminant and has a composition such that gas in equilibrium with it would have a concentration y_2' at the gas outlet, and y_1' at the gas inlet. The difference between y and y' at any point in the scrubber is the absorption driving force.

The counter-current scrubber has the highest potential removal efficiency, because it contacts the gas leaving the scrubber with the cleanest scrubbing liquor.

The concurrent scrubber does just the opposite and tends to bring the discharge gas into equilibrium with the most contaminated liquor. However, the co-current scrubbers are considerably less prone to plugging with solids than the counter-current, and also require less gas pressure drop to operate.

Both plugging potential and bringing about a close approach to equilibrium are important in the WPPA application, and a compromise between the efficiency of the counter-current and the mechanical advantages of the co-current scrubber is reached in the cross-flow packed scrubber illustrated by Figure 8c.

This type of scrubber has been widely used, either alone or in combination with spray towers, Venturis, or wet cyclonics in phosphoric acid plants. Figure 9 is a sketch of a cross-flow packed scrubber.(14) Combination scrubbers, treating gases from all the sources in the WPPA plant have also been used, as shown in Figure 10.(4)

In this discussion, the use of a cross-flow packed scrubber will be considered as the principal contacting device, either alone or preceded by a scrubber-cooler.

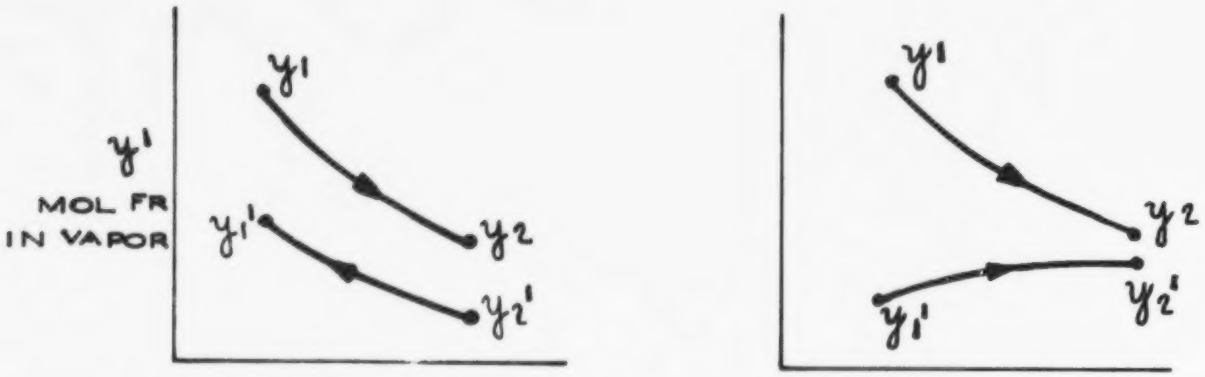
CHEMICAL REQUIREMENTS

The efficiency which can be obtained by a cross-flow scrubber is limited by the pond water composition if this is used alone as the scrubbing medium. In addition, the "pick-up" of fluorine as the water passes through the scrubber is of importance in setting the minimum concentration attainable.

For the case of a 500 T/D plant as illustrated in Table 10, the required outlet concentration is about 2.15 ppm F (this is approximate because the scrubber outlet gas flow will be slightly different than the inlet flow on which

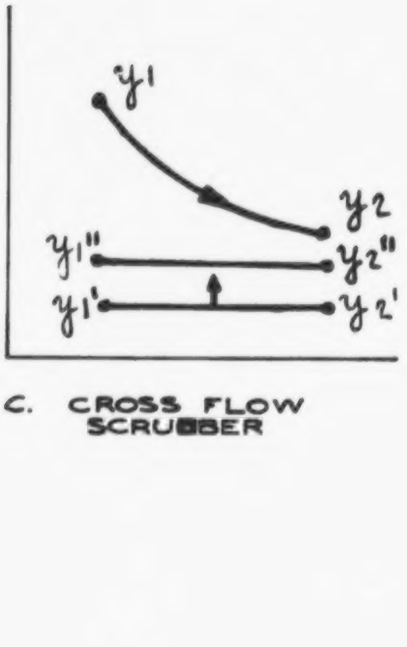
FIGURE 8

COMPARISON OF SCRUBBER TYPES

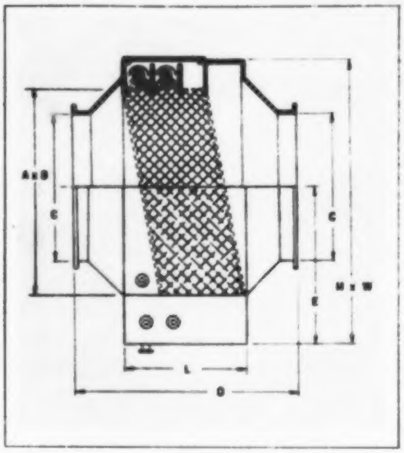

y1
y'
MOL FR
IN VAPOR
y1'
y2
y2'
y1
y2
y1'
y2'

a. COUNTER CURRENT SCRUBBER

b. CON-CURRENT SCRUBBER


y1
y2
y1''
y2''
y1'
y2'

c. CROSS FLOW SCRUBBER



CROSS-FLOW
PACKED SCRUBBER

Capacity ACFM T = 70°F	Liquid Rate GPM	Minimum Overall Length D	Overall Width W	Overall Height H	Base To C L Inlet E	Inlet Outlet Dimensions A x B	Duct Dia C	Shipping Weight (Lbs.)	Operating Weight (Lbs.)
2,100	8-16	6	[illegible]	4	1-9	2 x 2-6	12	525	1,100
4,000	11-22	7	[illegible]	5	[illegible]	3 x 3-6	16	800	1,600
5,900	14-28	7	3-6	5-6	[illegible]	3-6 x 4	20	1,250	2,100
8,000	17-34	7	[illegible]	6	[illegible]	4-3 x 4-6	24	1,275	2,600
10,000	19-38	7-6	[illegible]	6-6	[illegible]	4-9 x 5	26	1,550	3,100
12,000	21-42	7-6	[illegible]	[illegible]	[illegible]	5-3 x 5-6	30	1,800	3,500
13,800	22-44	8	5-6	[illegible]	[illegible]	5-6 x 6	32	2,050	3,900
16,300	24-48	8	6	9	[illegible]	6 x 6-6	36	2,350	4,400
18,000	25-50	8-6	[illegible]	[illegible]	4	6-3 x 7	36	2,600	4,800
20,500	26-52	9	[illegible]	9	[illegible]	6-6 x 7-6	40	2,900	5,250
22,000	28-56	9	7	9	[illegible]	7 x 7-6	40	3,100	5,600
24,400	29-58	9	[illegible]	[illegible]	4-6	7-3 x 8	42	3,300	6,000
26,300	30-60	9-6	7-6	[illegible]	4-8	7-6 x 8-3	42	3,600	6,400
28,500	31-62	10	7-9	[illegible]	4-10	7-9 x 8-9	44	3,900	6,900
30,000	32-64	9-6	8	10-6	5	8 x 9	48	4,000	7,100
32,000	33-66	10	8-3	[illegible]	[illegible]	8-3 x 9-3	48	4,200	7,500
34,000	36-72	9-6	9	10-6	5	9 x 9	48	4,400	7,900
36,000	36-72	10	9	11	[illegible]	9 x 9-6	48	4,700	8,300
38,000	38-76	9-6	9-6	11	[illegible]	9-6 x 9-6	54	4,850	8,600
40,000	38-76	10	9-6	11-6	5-6	9-6 x 10	54	5,100	8,950
42,000	40-80	10	10	11-6	6	10 x 10	54	5,400	9,350
44,000	40-80	10-6	10	12	[illegible]	10 x 10-6	54	5,650	9,750
46,300	42-84	10	10-6	[illegible]	[illegible]	10-6 x 10-6	60	5,800	10,100
48,500	42-84	[illegible]	[illegible]	[illegible]	6-6	10-6 x 11	60	6,100	10,500
50,000	43-86	[illegible]	[illegible]	[illegible]	6-6	[illegible] x 11	60	6,300	[illegible]

NOTE: 1. Minimum base length (L) is 4 ft. for all units.
2. All dimensions are approximate.
3. Pressure drop is approximately 0.4 in. W.C. per foot of packed depth.
4. Final dimensions will vary to meet specific operating conditions.

FIGURE 9

SKETCH OF A CROSS-FLOW PACKED SCRUBBER

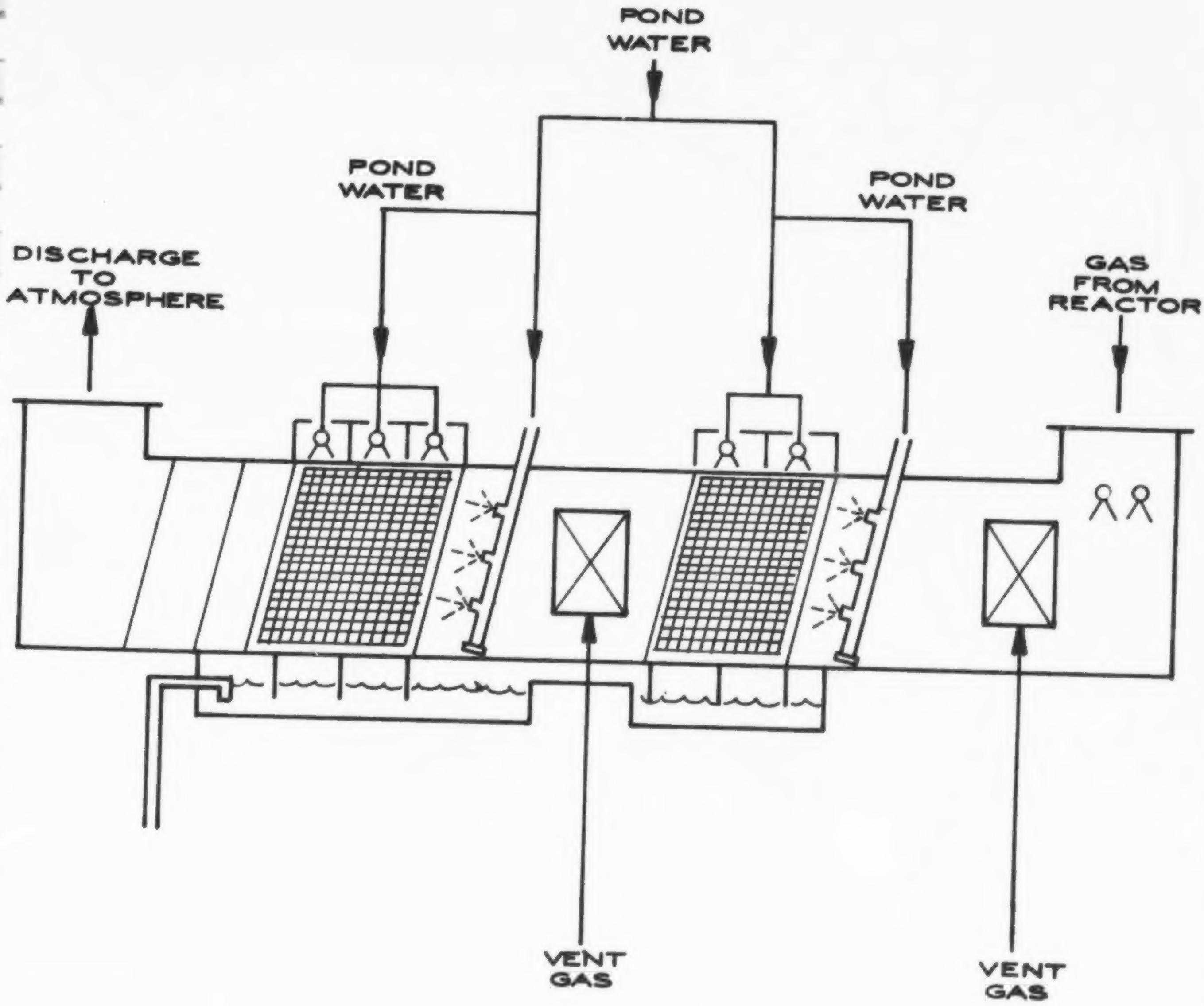

POND WATER
POND WATER
POND WATER
DISCHARGE TO ATMOSPHERE
GAS FROM REACTOR
VENT GAS
VENT GAS

FIGURE 10

COMBINATION SCRUBBER
FOR WPPA PROCESS

this value was based). In order to achieve low level, it is necessary that the water entering the scrubber have an equilibrium vapor pressure (y') lower than this value.

Table 13 lists a number of calculated concentrations in equilibrium with water of various fluosilicic acid contents and temperatures. A typical pond water is likely to contain between 0.2 and 1.5 wt.% H_2SiF_6 and average around 80° F. This is low enough to bring about acceptable scrubbing (provided the scrubber is large enough) as long as the concentration is below about 1.0 wt.% and the temperature in the scrubber is below 100° F. At 120° F, the pond water would have to be near 0.1 wt.% H_2SiF_6 for satisfactory performance.

The degree to which the scrubber approaches the limit set by the composition and temperature of the pond water is often measured in terms of the "Number of Transfer Units", or NTU. This is defined as

$$\text{NTU} = \int_{y_1}^{y_2} \frac{dy}{y-y'} \tag{14}$$

$$\text{NTU} = \ln \left[\frac{y_1 - y_1'}{y_2 - y_2'}\right] \tag{15}$$

where the scrubbing liquor is pure water, and the quantity of water flow is large, $Y_1' = y_2' = 0$ and

$$\text{NTU} = \ln \left[\frac{y_1}{y_2}\right] \tag{16}$$

A scrubber called upon to bring about a 95% reduction in fluorine concentration at a gas temperature of 100° F, operating in a range where the concentrations in equilibrium with the pond water would be negligible, and the NTU value required would be

$$\begin{aligned} \text{NTU} &= \ln \left[\frac{y_1}{0.05\ y_1}\right] \\ &= \ln\ 20 \\ &= 3.0 \end{aligned}$$

The operation of the cross-flow scrubber must be such as to reduce the concentration to the same range as that found in equilibrium with the pond water. This requires a great deal of additional contact as illustrated in Table 14.

TABLE 13

CONCENTRATION LIMITS AT SCRUBBER OUTLET

(Using H_2SiF_6 – Containing Pond Water)

ppm F

Temperature at Scrubber Discharge, °F	Fluorine* Content of Pond Water, Wt. %					
	0	0.1	0.5	1.0	1.5	2.0
80	0	0.4	0.7	0.9	1.1	1.3
100	0	1.0	1.45	1.8	2.3	2.6
120	0	2.1	2.7	3.5	4.3	5
140	0	3.5	4.8	6.1	7.6	9

*NOTE: H_2SiF_6 content is approximately $\frac{144}{114} = 1.26$ times fluorine content.

Here the NTU values required for reaching 0.16 lb/hr emission of F for a 500 T/D plant are given in terms of the pond water composition and temperature.

In preparing specifications for scrubbing systems, the two scrubbing efficiencies specified will, in effect, be those for 0.5 wt.% fluorine with and without a prescrubber.

TABLE 14

NTU REQUIRED TO REACH 2.15 ppm F(1)

DISCHARGE CONCENTRATION

(NTU Units)(2)

Temperature at Scrubber Discharge, °F	Fluorine Content of Water from Scrubber, Wt.%					
	0	0.11	0.5	1.0	1.5	2.0
80	6.2	6.28	6.59	6.72	6.91	7.22
100	6.2	6.72	7.32	8.00	NA	NA
120	6.2	10	NA	NA	NA	NA
140	6.2	NA	NA	NA	NA	NA

(1) Total fluorine as SiF_4 and HF, reported as F°

(2) Based on an inlet concentration of 1050 ppm F°.

SPECIFICATIONS AND COSTS

The preparation of specifications for scrubbing equipment was based on the definition of two hypothetical WPPA plants. The smaller of the two corresponds in size to the 500 T/day example used throughout this section. The larger process was taken as 900 T/day.

Ventilating air from the digester was presumed to be treated separately from the other effluent gases. This air was taken at 25,000 and 36,000 SCFM at the digester, and it was assumed that the gas was 75% saturated at 140° F. This would result in an adiabatic saturation temperature of about 132° F if insufficient cooling water were provided, and it would not be possible to achieve the required reduction in fluorine. Therefore, it is necessary for the scrubber manufacturer to take into account the additional water requirement to cool the gases to approximately 100° F in order to meet the efficiency requirements.

The ventilating air usage is not proportional to the plant throughput. For this reason, the larger process may be assumed to require a lower ventilating rate per ton of product than the smaller one. This is likely to pick up fluorine at about the same concentration level as the small unit and, therefore, have a potential for emission *per ton of product* which is lower than for a similarly designed small plant. However, for purposes of this specification, a scrubber with the same NTU, and presumably the same packing height, was specified so as to make scrubber gas flow the only variable between the two specifications.

In order to ascertain the variation in cost with scrubber size, a low efficiency case was specified in which the gas flow through the reactor was reduced by one-third, and the residual fluorine content held constant. This results in *increasing* the allowable concentration of fluorine in the effluent by a factor of three and reduces the efficiency requirement for the scrubber. In this case, the cost of the abatement system will be lower, even though the same emission level is expected.

In order to make a reasonable presentation of the cost pattern, it must be presumed that the cost is not fixed relative to plant size and total allowable emission, but rather varies with design gas flow rate and allowable emission. The freedom to reduce the ventilation rate through the digester allows the process designer to improve the emission potential of the system and at the same time to reduce the cost of pollution abatement equipment.

TABLE 15

PROCESS DESCRIPTION FOR WET PROCESS PHOSPHORIC ACID PLANT CROSS-FLOW SCRUBBER

This specification describes a scrubbing system to serve a Prayon-designed WPPA facility of the specified size. The scrubbing system is to comprise a cross-flow packed scrubber of non-plugging design, a fan to overcome the pressure drop through the scrubber and ductwork, a recirculating pump (if required), a 100 ft. free-standing stack, and all interconnecting ductwork. The scrubber supplier is to furnish the ductwork connecting the digester to the scrubber, which will be approximately 120 feet long.

The scrubbing system is intended to serve only the reactor air vent. Other arrangements will be provided for the filter ventilation and sump vents.

Pond water is available at 80°F with the following properties:

	Design	*Min.*	*Max.*
Pond Water pH	2.0	1.2	2.2
Temp., °F	80.0	55	88
SO_4, *wt. %*	0.15	–	–
P_2O_5, *wt. %*	0.1	–	–
H_2SiF_6, *wt. %*	0.63	0.25	1.0
Fluorine, wt. %	0.5	0.2	0.8

The scrubber is required to produce the design performance when operating with the "design" pond water conditions. The scrubber manufacturer shall specify the required water circulation rate through the scrubber in order to accomplish both cooling of the reactor effluent gases to 100°F or lower and absorption of fluorine containing gases at the specified level.

Materials of construction shall be limited to the following, which shall be selected for hydrofluoric acid service:

PVC

Rubber (below 160° F)

FRP (Dynel lined)

No metal parts shall be used where exposure to process gas or pond water may be significant.

TABLE 16

OPERATING CONDITIONS FOR WET PROCESS PHOSPHORIC ACID PLANT CROSS-FLOW PACKED SCRUBBER

	Small	*Large*
Plant Capacity, Ton/Day P_2O_5	*500*	*900*
Acid Strength, Wt. % P_2O_5		
From Digesters	*30*	*30*
From Evaporators	*54*	*54*
Fluorine Content, Wt. % F	*1.5*	*1.5*

	A	*B*	*C*	*D*
Gas to Scrubbers				
Flow, SCFM	*8,333*	*25,000*	*12,000*	*36,000*
Flow, DSCFM	*7,040*	*21,050*	*10,100*	*30,300*
Flow, ACFM	*9,433*	*28,300*	*13,533*	*40,600*
Temp., °F	*140*	*140*	*140*	*140*
Moisture, Vol. %	*15.7*	*15.7*	*15.7*	*15.7*
Fluorine(1), lb/hr	*60*	*80*	*86.6*	*115*
Fluorine(1), ppm	*2,400*	*1,050*	*2,400*	*1,050*
Particulate, lb/hr	*3.6*	*10.8*	*5.2*	*15.5*
Particulate, gr/SCF	*0.05*	*0.05*	*0.05*	*0.05*
Gas from Scrubbers				
Flow, SCFM	*7,540*	*22,600*	*10,800*	*32,500*
Flow, DSCFM	*2,040*	*21,050*	*10,100*	*30,300*
Flow, ACFM	*7,960*	*23,700*	*11,000*	*34,400*
Temp. ° F	*100*	*100*	*100*	*100*
Moisture, Vol. %	*6.7*	*6.7*	*6.7*	*6.7*
Fluorine, lb/hr	*0.16*	*0.16*	*0.29*	*0.29*
Fluorine, ppm	*7.1*	*2.35*	*8.9*	*2.95*
Fluorine Removal, Wt. %	*99.73*	*99.80*	*99.66*	*99.75*

(1) *Total Fluorine reported as F^-.*

Particulate, lb/hr	0.032	0.097	0.46	1.4
Particulate, gr/SCF	0.005	0.005	0.005	0.005
Particulate Removal, Wt. %(5)	91.1	91.1	91.1	91.1
Estimated y', ppm	1.4	1.4	1.4	1.4
Estimated NTU required	5.92(3)	7.05(4)	5.75(3)	6.50(2, 4)

(1) *It is not expected that this particulate removal specification will influence the scrubber design.*

(2) *Please quote a scrubber with the same packing depth and NTU as for the "small" unit – i.e. 7.05 NTU.*

(3) *These should be quoted on the basis of a rounded off NTU value of 6. Please see the supplementary scrubber price sheet attached.*

(4) *These should be quoted on the basis of a rounded off NTU value of 7.5.*

(5) *Total fluorine reported as F^o.*

TABLE 17

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR WET SCRUBBERS FOR WPPA PROCESS PLANTS

	One-third Gas Flow		Normal Gas Flow	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	9,433	13,533	28,300	40,600
°F	140	140	140	140
SCFM	8,333	12,000	25,000	36,000
Moisture Content, Vol. %	15.7	15.7	15.7	15.7
Effluent Contaminant Loading				
(ppm), Fluorine	2,400	2,400	1,050	1,050
lb/hr, Fluorine	60	86.6	80	115
Cleaned Gas Flow				
ACFM	7,960	11,000	23,700	34,400
°F	100	100	100	100
SCFM	7,540	10,800	22,600	32,500
Moisture Content, Vol. %	6.7	6.7	6.7	6.7
Effluent Contaminant Loading				
(ppm), Fluorine	7.1	8.9	2.35	2.95
lb/hr, Fluorine	0.16	0.29	0.16	0.29
Cleaning Efficiency, %	99.73	99.66	99.66	99.75
(1) Gas Cleaning Device Cost	8,370	9,685	17,703	21,576
(2) Auxiliaries Cost	4,357	5,223	8,533	9,473
(a) Fan(s)	2,240	3,066	4,886	5,993
(b) Pump(s)	2,037	2,067	3,495	3,310
(c) Damper(s)	--	--	--	--
(d) Conditioning, Equipment	--	--	--	--
(e) Dust Disposal Equipment	--	--	--	--
(3) Installation Cost	25,832	28,450	36,270	40,865
(a) Engineering	1,500	1,750	2,000	2,250
(b) Foundations & Support	2,355	2,417	3,000	3,125
(c) Ductwork	--	--	--	--
(d) Stack	4,812	5,140	7,285	8,025
(e) Electrical	--	--	--	--
(f) Piping	3,300	3,750	4,575	4,925
(g) Insulation	--	--	--	--
(h) Painting	--	--	--	--
(i) Supervision	--	--	--	--
(j) Startup	710	710	710	710
(k) Performance Test	1,475	1,475	1,475	1,475
(l) Other	6,900	8,000	9,150	10,250
(4) Total Cost	38,559	43,358	62,506	71,914

FIGURE 11

CAPITAL COSTS FOR WET SCRUBBERS FOR WPPA PROCESS PLANTS

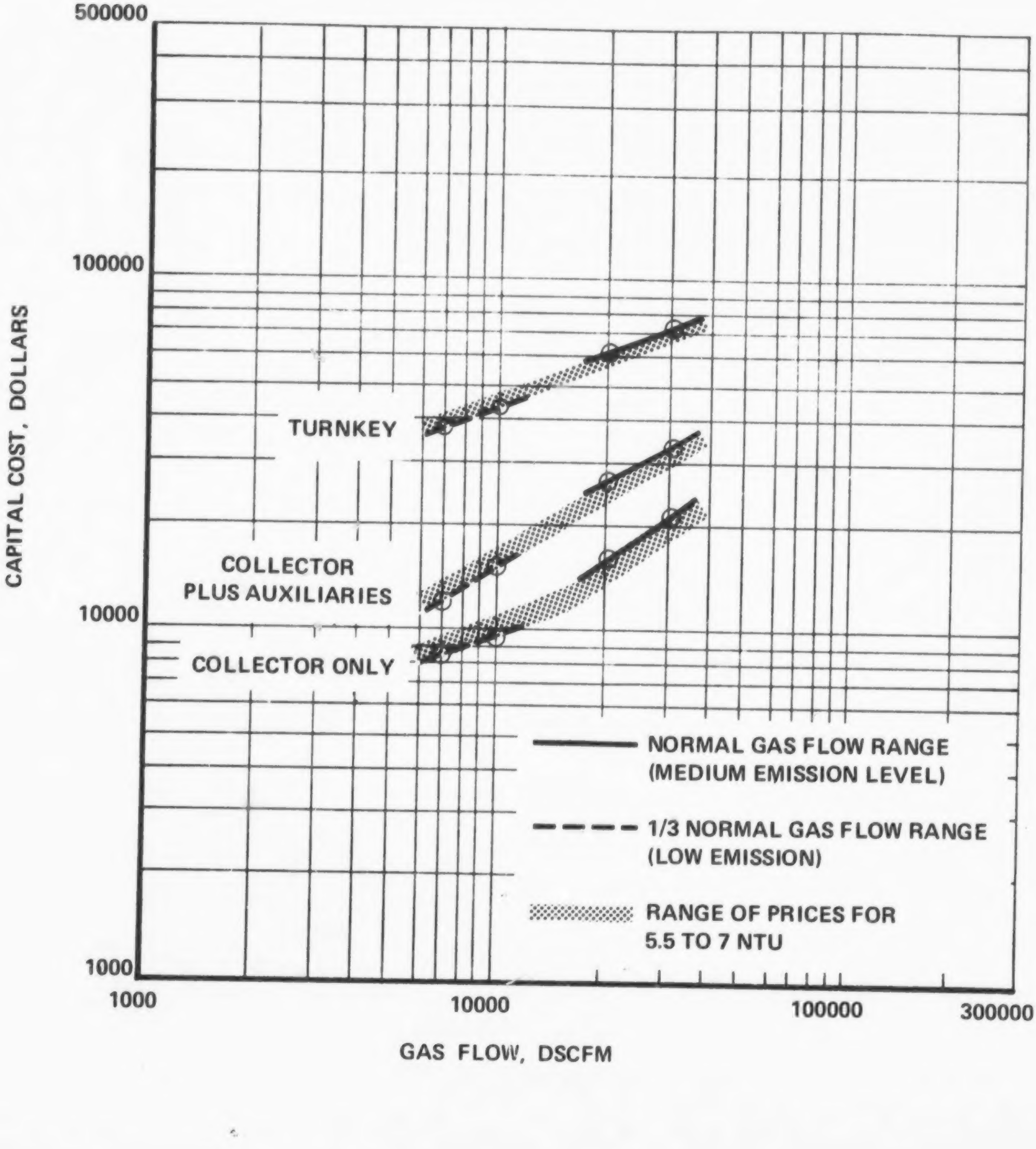


TABLE 18

CONFIDENCE LIMITS FOR CAPITAL COST OF SCRUBBERS FOR WPPA PROCESS PLANTS

Population Size – 5

Sample Size – 2

Capital Cost = $17,702

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$16,663	$18,741
75	15,401	20,004
90	12,930	22,474
95	10,262	25,142

Capital Cost = $21,576

Conf. Level, %	Lower Limit	Upper Limit
50	$20,559	$22,593
75	19,323	23,828
90	16,905	26,246
95	14,293	28,858

FIGURE 12

CONFIDENCE LIMITS FOR CAPITAL COST OF SCRUBBERS FOR WPPA PROCESS PLANTS

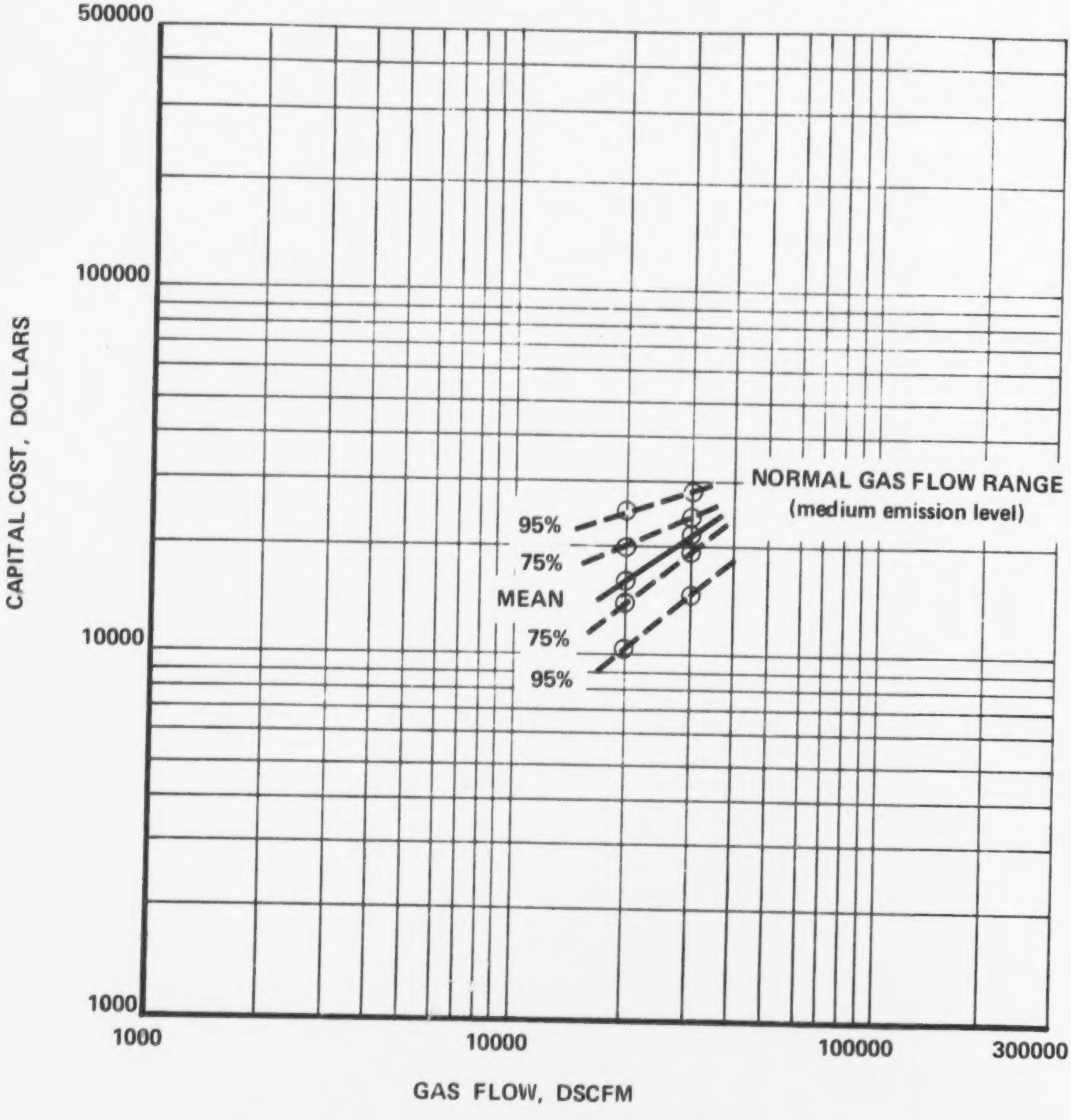


TABLE 19

CONFIDENCE LIMITS FOR CAPITAL COST OF INSTALLED SCRUBBERS FOR WPPA PROCESS PLANTS

Population Size – 5 Sample Size – 2

Installed Cost = $62,506

Conf. Level, %	Installed Cost, Dollars Lower Limit	Upper Limit
50	$59,227	$65,785
75	55,244	69,767
90	47,449	77,563
95	39,029	85,982

Installed Cost = $71,914

Conf. Level, %	Lower Limit	Upper Limit
50	$71,609	$72,219
75	71,238	75,589
90	70,513	73,314
95	69,730	74,097

FIGURE 13

CONFIDENCE LIMITS FOR CAPITAL COST OF INSTALLED SCRUBBERS FOR WPPA PROCESS PLANTS

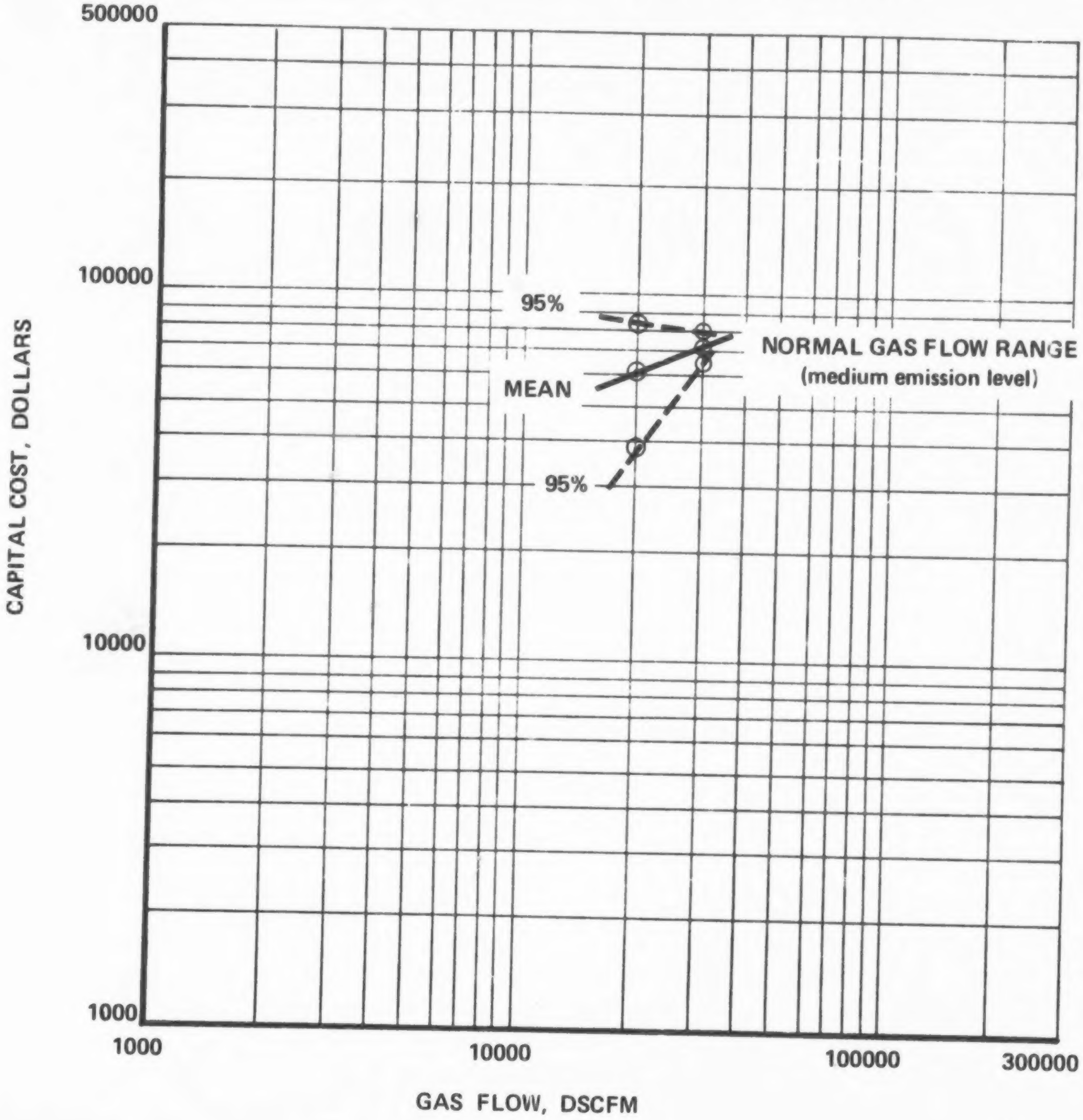


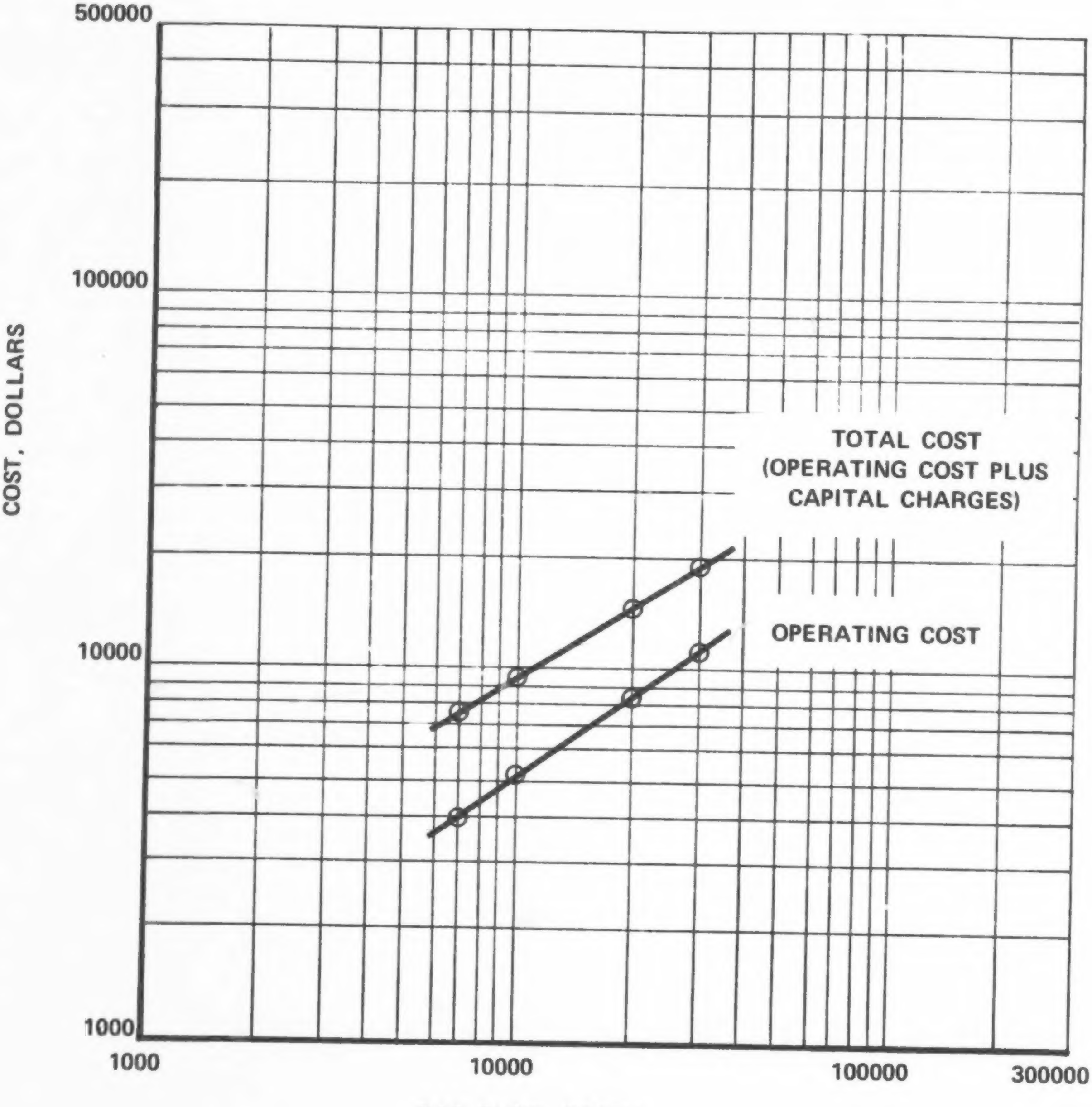
TABLE 20

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR WET SCRUBBERS FOR WPPA PROCESS PLANTS

Operating Cost Item	Unit Cost	One-third Gas Flow		Normal Gas Flow	
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor		$ 854	$1,504	$ 2,404	$ 4,204
Maintenance					
Labor					
Materials					
Total Maintenance		1,920	2,200	3,105	3,415
Replacement Parts					
Total Replacement Parts		75	100	150	200
Utilities					
Electric Power	$.011/kw-hr	850	1,400	2,770	4,350
Fuel					
Water (Process)	$.25/M gal				
Water (Cooling)	$.05/M gal				
Chemicals, Specify					
Total Utilities		850	1,400	2,770	4,350
Total Direct Cost		3,703	5,209	8,387	12,182
Annualized Capital Charges		3,856	4,336	6,251	7,191
Total Annual Cost		7,559	9,545	14,638	19,373

FIGURE 14

ANNUAL COST FOR WET SCRUBBERS
FOR WPPA PROCESS PLANTS



SUPERPHOSPHORIC ACID

Superphosphoric acid (SPA) is a product of further dehydration of 54% P_2O_5 phosphoric acid. It is useful in the manufacture of high quality fertilizers, and in particular liquid fertilizers.

Because the further dehydration of phosphoric acid drives off most of the remaining fluorine, super phosphoric acid is also prepared as an intermediate in the production of low-fluorine content phosphoric acid or fertilizers.

Ordinarily, superphosphoric acid is prepared at concentrations in the range of 72% P_2O_5 which corresponds to approximately 100% H_3PO_4. However, the product is not the pure orthophosphoric acid but a complex mixture of ortho, pyro, tri and meta phosphoric acid forms.(2)

PROCESS DESCRIPTION

Phosphoric acid of about 54 wt.% P_2O_5 is charged to the process as shown in Figure 15. Depending on the pretreatment of the 54% acid prior to storage, and on the intended use of the SPA, further centrifugation or settling for the removal of solids may be employed. The presence of solids in the SPA is considerably more objectionable than in the ortho acid, especially if SPA is to be used in the manufacture of liquid fertilizers.(6)

The 54% acid is pumped into the vacuum flash evaporator circulating stream before the heat exchanger and admixed with hot circulating acid. The mixture is circulated by gravity, using the density differential of the boiling acid between the riser and the relatively cool acid in the return leg as the driving force. The partially vaporized mixture is discharged into the vacuum flash vessel where steam, SiF_4 and HF are disengaged from the concentrated acid. The liquid phase collects in the flash vessel and flows by gravity down the return leg. Part of the liquid is drawn off at this point to the product cooling tank and part is recycled to the heat exchanger.

The method of heating and vaporizing varies considerably from one process design to another. The vacuum flash evaporator may be replaced by a falling-film evaporator, in which the heat transfer and phase separation are accomplished in a single vessel, or by a non-recirculated single-stage flash process. The configuration of the vaporizer is not as important as the down-stream equipment from an air pollution control standpoint.(6)

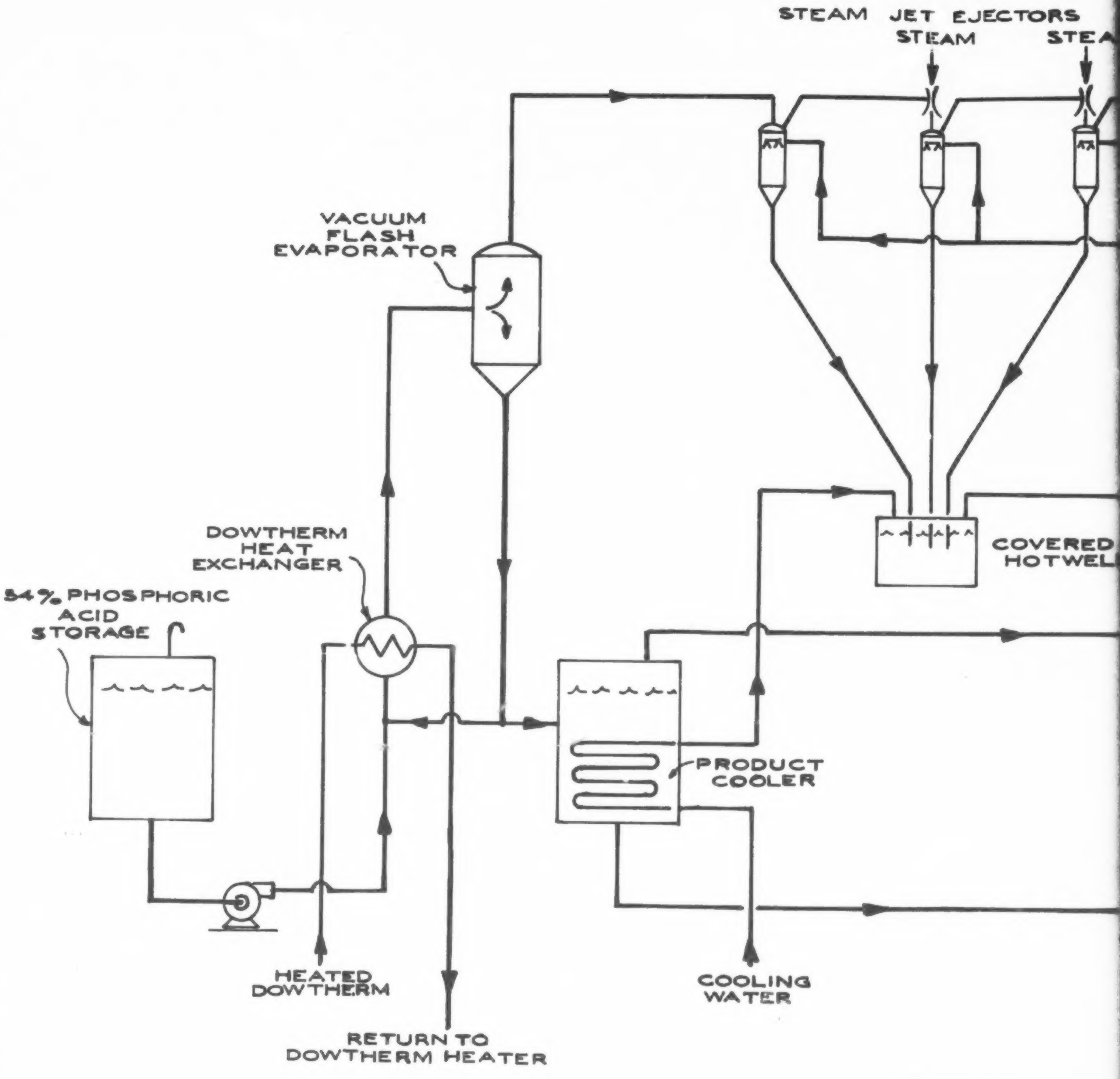

STEAM JET EJECTORS
STEAM
STEA
VACUUM FLASH EVAPORATOR
DOWTHERM HEAT EXCHANGER
54% PHOSPHORIC ACID STORAGE
COVERED HOTWEL
PRODUCT COOLER
HEATED DOWTHERM
RETURN TO DOWTHERM HEATER
COOLING WATER

FIGURE 15

SCHEMATIC FLOW DIAGRAM FOR

VACUUM FLASH SUPERPHOSPHORIC ACID

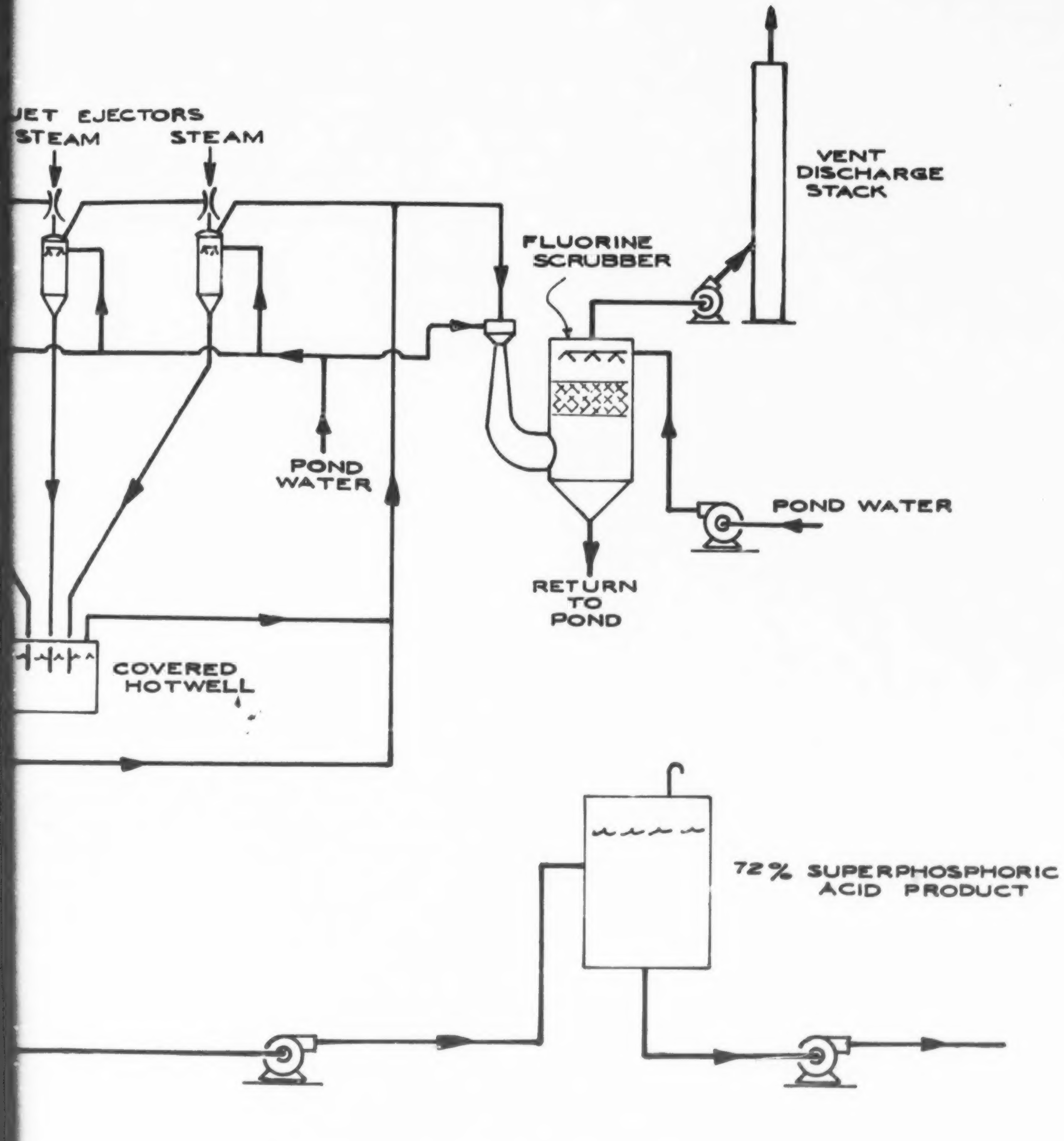

JET EJECTORS
STEAM
STEAM
VENT
DISCHARGE
STACK
FLUORINE
SCRUBBER
POND
WATER
POND WATER
RETURN
TO
POND
COVERED
HOTWELL
72% SUPERPHOSPHORIC
ACID PRODUCT

FIGURE 15

FLOW DIAGRAM FOR

ERPHOSPHORIC ACID PROCESS

In several existing plants, the vaporization is accomplished by direct contact of hot combustion gases with the phosphoric acid in an evaporation chamber. This process, described as the "submerged combustion process" differed widely from that described here. One of the principal draw-backs of the process was the substantial volume of effluent gases contaminated with fluorine. This problem has contributed to the reduced popularity of the submerged combustion process.

The steam, acid fumes and non-condensable gases discharged from the vacuum flash evaporator are cooled in a barometric condenser by circulating pond water, and most of the steam and fluorine-containing compounds are removed from the barometric with the water for return to the pond. It is probable that the principal form of the fluorine compounds at the barometric condenser inlet is HF, and that contact with the pond water causes an equilibration between the HF and SiF_4, which is the principal gaseous fluorine compound leaving with the barometric condenser gaseous effluent.(8)

The barometric condenser is likely to operate at a high temperature – 130-160°F and the fluorine content of the steam-acid vapors discharged into the condenser will be on the order of:

$$\frac{\text{wt. F}}{\text{wt. } H_2O} = \frac{\text{Fluorine removed from acid}}{\text{Water removed from acid}}$$

$$\frac{\text{wt. F}}{\text{wt. } H_2O} = \frac{1 \cdot (0.015 - 0.003)}{1 \cdot (0.25 - 0.00)^*} = \frac{0.012}{0.25} = 0.048 \frac{\text{lb F}}{\text{lb } H_2O}$$

or nearly 5% fluorine. Condensation of this material in a surface condenser would produce high concentrations of fluorine in the non-condensable gas phase. However, the use of a large volume of pond water to condense the steam provides for dilution of the liquid-phase fluorine content to a level closer to that of the pond water than the calculated condensate concentration. The concentration of fluorine discharged from the barometrics can be presumed to be something in excess of the concentration in equilibrium with pond water at the barometric condenser operating temperature.

*Water removed from acid in SPA process is based on:

54% P_2O_5 = 74.5% acid, 25.5% water

72% P_2O_5 = 100% acid, 0% water

Preceding page blank

One or two stages of steam-jet ejectors may be used to pump non-condensable gases and residual steam out of the barometric condenser and up to atmospheric pressure. Each steam-jet ejector is equipped with a barometric condenser to condense the steam used for motive power and the residual steam present at the inlet.

In the flow scheme illustrated in Figure 15, it is apparent that the non-condensables entering the system, steam load from the final barometric, and the fluorine compounds which contaminate the combination of these two constitute the only process emission source in the SPA plant. Two other sources are usually piped into the same emission control device as the barometric condenser vent. These are:

(1) the hotwells into which the discharge of cooling water from the barometrics and product coolers flows, and from which pumps for return of the cooling water to the gypsum pond take suction.

(2) the cooling tank.

Neither of these can be considered a process source in the ordinary use of the term, because there is no gas generation or air usage required by the process. In the case of the hotwells, it is a matter of convenience that these are allowed to "breathe" through relatively loose-fitting covers which eliminate the need for pressure controls to prevent flashing of water at the return pump suction, and allow for easy access for cleanout and maintenance of the hotwell. Air leakage through the covers is likely to provide the main gas source to the pollution abatement device.

The hot liquid effluent from the flash evaporator is likely to be cooled in a tank with cooling coils or surfaces in it. This is because the scaling properties of the acid make it difficult to use shell-and-tube coolers. A cooler-surge tank of this type requires a vent to accommodate changes in the liquid level, even though the gas flow entering the tank on decreases in level ought to just equal the gas flow leaving as the level increases. A vapor-tight surge system could provide for intermittant breathing of the tank without any net gas flow. However, such a system would be costly to build and maintain, and would offer little advantage over simply venting the tank into the inlet of the condenser vent scrubber.

The air pollution control device shown in Figure 15 as a Venturi scrubber followed by a packed scrubbing section which serves as a gas absorber and mist eliminator, uses pond water to provide final cleanup of the non-condensable gases and air leakage into the system before discharge to the atmosphere. This gas stream is likely to be quite small, and can be treated by any of the scrubbing devices which can handle small volumes of gas and do an effective job of collecting aerosols and gaseous fluorides. A water-induced Venturi, or a conventional Venturi-fan combination as shown, are the most commonly used.

CHEMISTRY OF THE PROCESS

"Phosphoric acid" may be considered as a dilute form of orthophosphoric acid:(2)

$$\begin{array}{ccccc} 392\text{ lb} & = & 284\text{ lb} & + & 108\text{ lb} \\ 4H_3PO_4 & = & P_4O_{10} & \cdot & 6H_2O. \end{array} \tag{17}$$

At 54% P_2O_5 by weight, the concentration of orthophosphoric acid in water solution is $54 \times \frac{392}{284} = 74.5$. That is, a 100 pound sample of 54% P_2O_5 acid may be assumed to contain 74.5 pounds of orthophosphoric acid plus 25.5 pounds of water.

More concentrated acids can be produced either by contacting the ordinary phosphoric acid with anhydrous P_2O_5 or by evaporation of some of the water. In either case, removal of the free water alone would limit the concentration to 72.5% P_2O_5 which corresponds to 100% H_3PO_4.

It is possible to further dehydrate the orthophosphoric acid by removal of some of the molecular water, which converts the acid to forms other than ortho. The forms known to exist(2) are listed in Table 21. The ions formed by these acids are shown schematically in Figure 16.

Actually the dehydrated ortho acid species are mixed at most concentrations above that for pure ortho, and form increasingly viscous solutions as water is removed. Although vaporization of water from concentrated acid can produce semi-solid syrupy products, pure solid forms of the acids cannot be obtained by this procedure. Pure ortho and phosphoric acids have been prepared by chemical synthesis. Orthophosphoric acid is a white solid with a melting point of 42.3°C, and orthophosphoric acid hemi-hydrate, $2H_3PO_4 \cdot H_2O$, melts at 29.35°C. Crystalline pyrophosphoric acid melts at about 61°C.(2)

In order to obtain the concentrated polyacid forms, it is necessary to drive

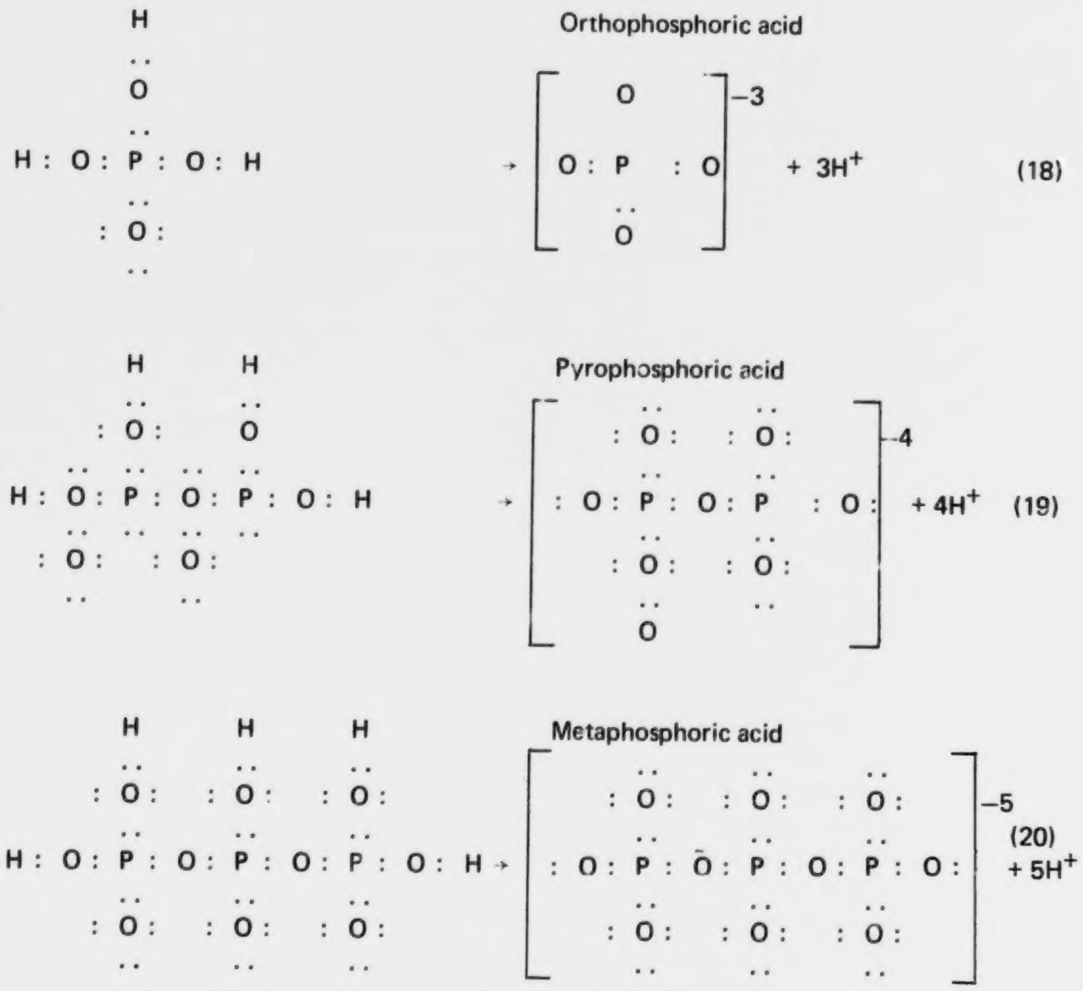
TABLE 21

FORMS OF PHOSPHORIC ACID

NAME	FORMULA	"% P_2O_5"
Orthophosphoric or Phosphoric Acid	$P_4H_{10} : 6H_2O$ or H_3PO_4	74.5
Pyrophosphoric	$P_4H_{10} : 4H_2O$ or $H_4P_2O_7$	80
Triphosphoric or Tripolysphosphoric Acid	$P_4H_{10} : \frac{10}{3} H_2O$ or $H_5P_3O_{10}$	82.5
Metaphosphoric Acid or Tetrapolyphosphoric Acid	$P_4H_{10} : 2H_2O$ or HPO_3	88.7

FIGURE 16

SCHEMATIC REPRESENTATION OF POLYPHOSPHORIC ACID FORMS


Orthophosphoric acid
−3
+ 3H+
(18)
Pyrophosphoric acid
−4
+ 4H+
(19)
Metaphosphoric acid
−5
(20)
+ 5H+

off most of the H_2SiF_6 present in the 54% acid. This gives rise to some potential for air pollution, although most of the fluorine-containing species are automatically condensed with the water driven from the acid and recondensed in barometric condensers. The fact that about 80% of the fluorine is distilled out of the acid during the concentration process leads to the use of superphosphoric acid to make low-fluorine orthophosphoric. As the super acid is diluted, it reverts to the ortho form spontaneously.

In the vaporization process, temperatures as high as 700-800°F are used.(6) The basic process is a simple flash distillation from which a vapor consisting of water, SiF_4, HF and traces of phosphoric acid is discharged, and a liquid consisting of mixed phosphoric acids, including:

orthophosphoric
pyro
polyphosphoric
tripolyphosphoric
tetrapolyphosphoric
H_2SiF_6
Impurities

is produced.

POLLUTION CONTROL CONSIDERATIONS

The gas stream leaving the barometric condensers should consist mainly of the air absorbed into the phosphoric acid charge plus air leakage into the vacuum section of the process. While the absorbed air flow rate can be calculated, the leakage rate cannot, as it depends upon the physical state of the equipment. For this reason it is customary to base the design of the steam jet ejectors on an assumed maximum leakage rate, and then to maintain the system so as to stay within the design rate. Typical design rates for SPA plants are:

Plant Capacity T/D P_2O_5	Non-condensable SCFM
100	150
300	200
500	250

The exit gas temperature from the final condenser stage is set by the ratio of cooling water to steam used. Figure 17 is a plot of a temperature vs. ratio for a particular design. Thus for a non-condensable gas rate of 200 SCFM and a ratio of 15/1 cooling water/steam on the last condenser stage, the vent from the final condenser will have a composition as follows:

	SCFM	M.W.	lb/hr	mol %
non-condensables	200	28.8	942	75.8
water vapor	64	18	182	24.2
HF	–	20	.00083	1 ppm
SiF_4	–	104	.0085	2 ppm
Total fluorine	–	–	.008	~ 10 ppm
	264	24.5	1,024	

For other ratios of pond water to steam and other non-condensable gas rates, similar tables could be constructed for purposes of specifying the inlet to a scrubber for removal of fluorine.

The hotwell and tank vents are also combinations of air and steam-HF/SiF_4 vapors, where the total volume of gas is set by the design of the hotwell covers and the composition and quantity of contaminants is determined by the outlet temperature from the barometrics. In this case, it is the average temperature of the water used in all of the barometric condensers (if a combined hotwell is used) rather than the temperature from the last stage.

The leakage rate may be set by designing the covers to limit the area of cracks around access ways and other openings, and then by designing the ventilating system to maintain a slight negative pressure inside the enclosure to induce an inward leakage velocity on the order of 200 FPM. This requires a negative pressure on the order of:

$$\begin{aligned} \Delta P &= \frac{\rho V^2}{2g} \\ &= \frac{0.05 \times \left(\frac{200}{60}\right)^2}{2 \times 32.2} = .0086 \text{ lb/ft}^2 \qquad (21) \\ &= 0.0017 \times \frac{62.4}{12} = 0.0045 \text{ inches w.c.} \end{aligned}$$

A hotwell eight feet square with a 1/4 in. crack all around the edge would have an area of:

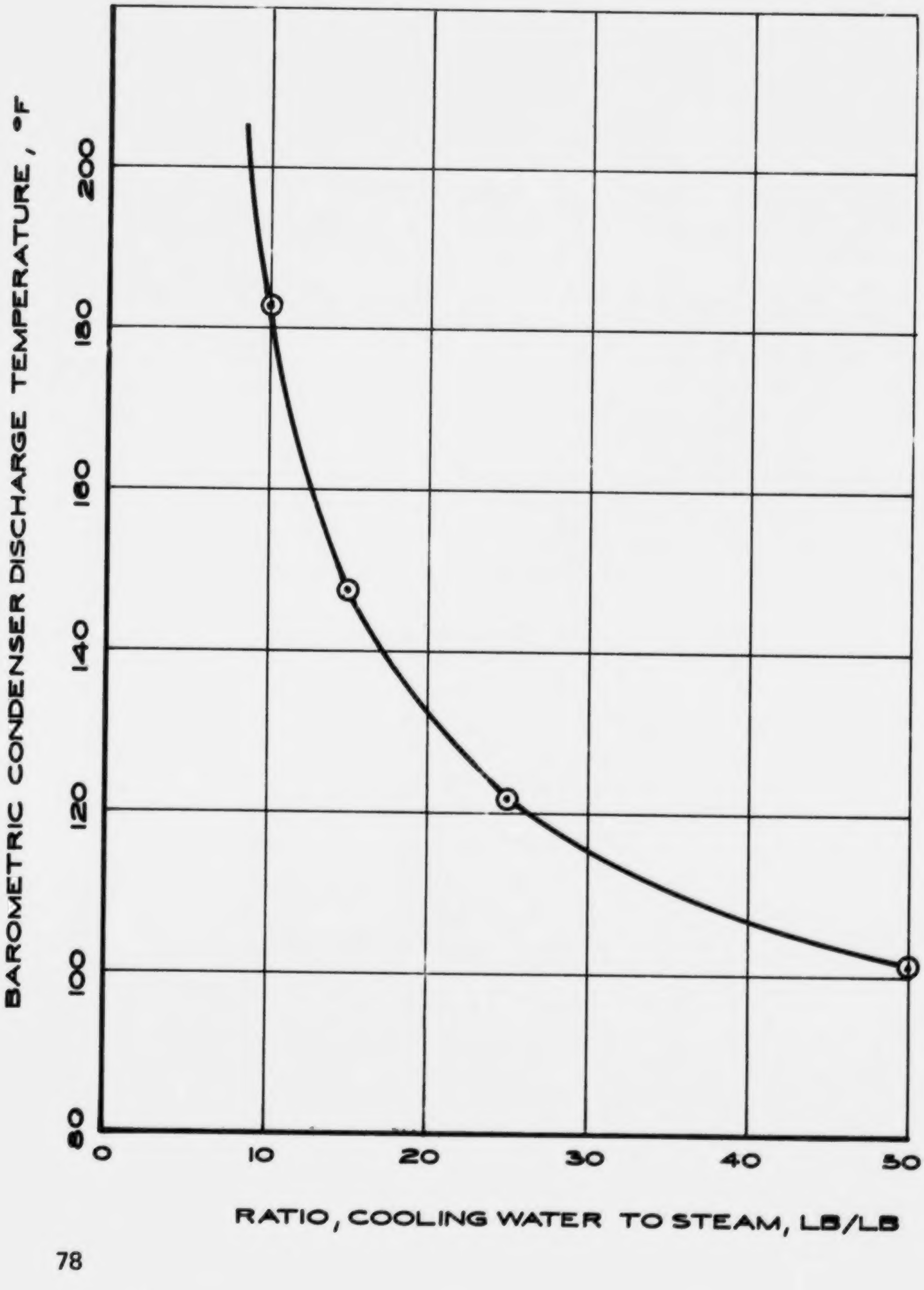
$$(8 + 8 + 8 + 8) \times \frac{1/4}{12} = 2/3 \text{ ft}^2$$

FIGURE 17

CONDENSER TEMPERATURE

VS. COOLING WATER/STEAM RATIO

(FOR 80°F POND WATER)



This leakage rate is likely to be overshadowed by the need for an open access way for maintenance and inspection. This would require about 250 FPM ventilation rate and might be as large as 2' x 4', which would result in 2000 SCFM inward leakage. This gas flow is likely to comprise the main gas emission source from the process. It is relatively arbitrary, in that the designer of the process can limit the flow rate to any degree he chooses by careful design of the enclosure. It is common to install a scrubber of nominal capacity, on the order of 2500 SCFM for this service, regardless of the size of the plant.

Particulate contaminants are limited to the liquid aerosols produced by condensation within the process, or by mists formed by mechanical action in the evaporator, condensers or elsewhere. The direct contact heat exchange which brings about condensation of the water in the barometric condensers is not likely to produce fine particle-size aerosols. The mechanically produced mists have a large particle size, and can be removed from the gas stream at very high efficiency by simple wet scrubbers. The principal problem is that of removal of gaseous fluorides and a satisfactory scrubber for gaseous fluoride removal will ordinarily be adequate for removal of the particulate mists.

In the case where falling film evaporators, spray towers, or other specially designed evaporators are used in place of the vacuum flash evaporator, it may be possible to generate sub-micron aerosols in the evaporator. For this reason, it should not be assumed that particulate emissions can be ignored if an unusual evaporator is incorporated into the system.

GAS CLEANING EQUIPMENT

As in the wet process for phosphoric acid manufacture, the SPA process emissions are treated exclusively by wet scrubbers. This is due to the need for gaseous fluorine removal from a wet gas stream, and the ready availability of scrubbing liquor from the gypsum pond.

Three significant factors contribute to the frequent selection of water-induced Venturi scrubbers for this service:

1. The gas flow is very low during normal operation.

2. There is likely to be a substantial variation in gas flow with time.

3. The water-induced scrubber does not require a fan.

The water-induced Venturi scrubber, shown schematically in Figure 18, is well suited for small gas flows, but seldom used for gas volumes over 5,000 ACFM. This is because it uses large quantities of water – relative to the gas flow – to pump the gas through the scrubber, the inlet ductwork and the discharge stack. Where the flow is not large, the inefficiency involved in the use of the water jet is of less importance than the mechanical simplicity and reliability gained by elimination of a fan from the system.

A second advantage of the water-induced Venturi for this service lies in its relative insensitivity to reductions in gas flow rate.

Packed scrubbers and conventional Venturi scrubbers rely upon the velocity of the gas passing through the scrubber for the energy required to give good contacting of gas and liquid. For this reason they are likely to show decreasing efficiency if the gas flow drops below 50 or 75 percent of the design flow rate. The water-induced Venturi does not rely on the gas flow for motive power, and will maintain high efficiency levels at extremely low gas flow rates, provided the water supply is not reduced. This is particularly important where the scrubber is arranged to handle gas leakage into a system with an access door which may be opened periodically.

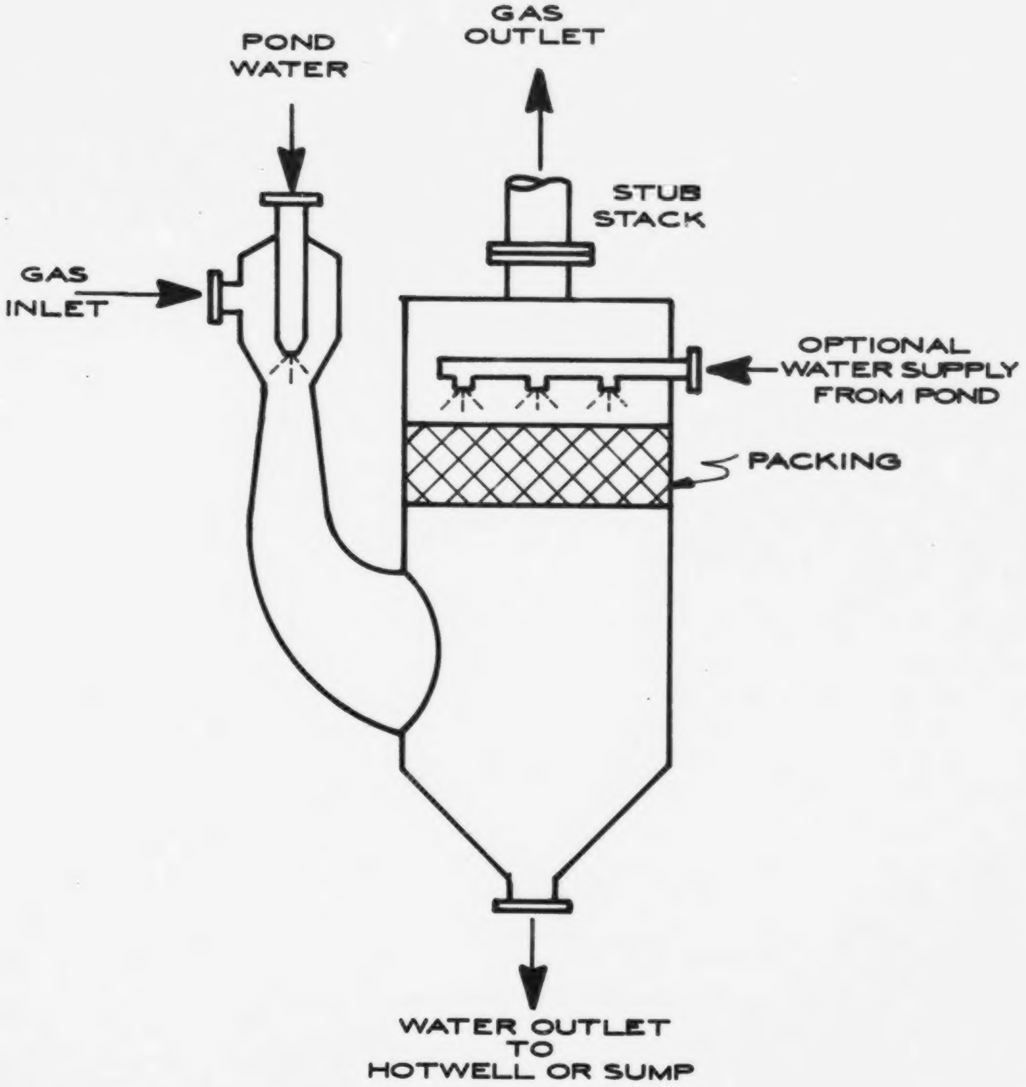
The water-induced Venturi is concurrent and limited to one theoretical stage and therefore it may be necessary to add a packed section in the entrainment separator vessel which serves to remove liquid droplets which become entrained in the throat as well as to provide for additional gas absorption when the gas flow rates are the highest.

The scrubbing requirements for SPA plants are rather nominal, and other types of scrubbers, such as cross-flow packed units, conventional Venturi scrubbers, mobile packed scrubbers and spray towers, each with an appropriate fan, are probably acceptable but mechanically more complicated alternatives to the water-induced Venturi in this application. A conventional Venturi scrubber with a packed gas absorption section in the entrainment separator is shown in Figure 19.

FIGURE 18

SCHEMATIC DRAWING OF

WATER-INDUCED VENTURI SCRUBBER


GAS OUTLET
POND WATER
STUB STACK
GAS INLET
OPTIONAL WATER SUPPLY FROM POND
PACKING
WATER OUTLET TO HOTWELL OR SUMP

SPECIFICATIONS AND COSTS

Specifications were written for a scrubber for this application on the assumption that the plant size does not significantly affect the size of the scrubber. Rather, the requirement that the air leakage into the hot wells into which the barometric condensers discharge is the main source of gas, and the leakage into the hot wells is more a function of the type of access-way ventilation required than the size of the plant.

Similarly, it was assumed that the emissions from hot wells and vents were nominal, and that a nominal scrubber efficiency would be satisfactory.

FIGURE 19

SCHEMATIC DRAWING OF

VENTURI SCRUBBER WITH

PACKED ENTRAINMENT SEPARATOR

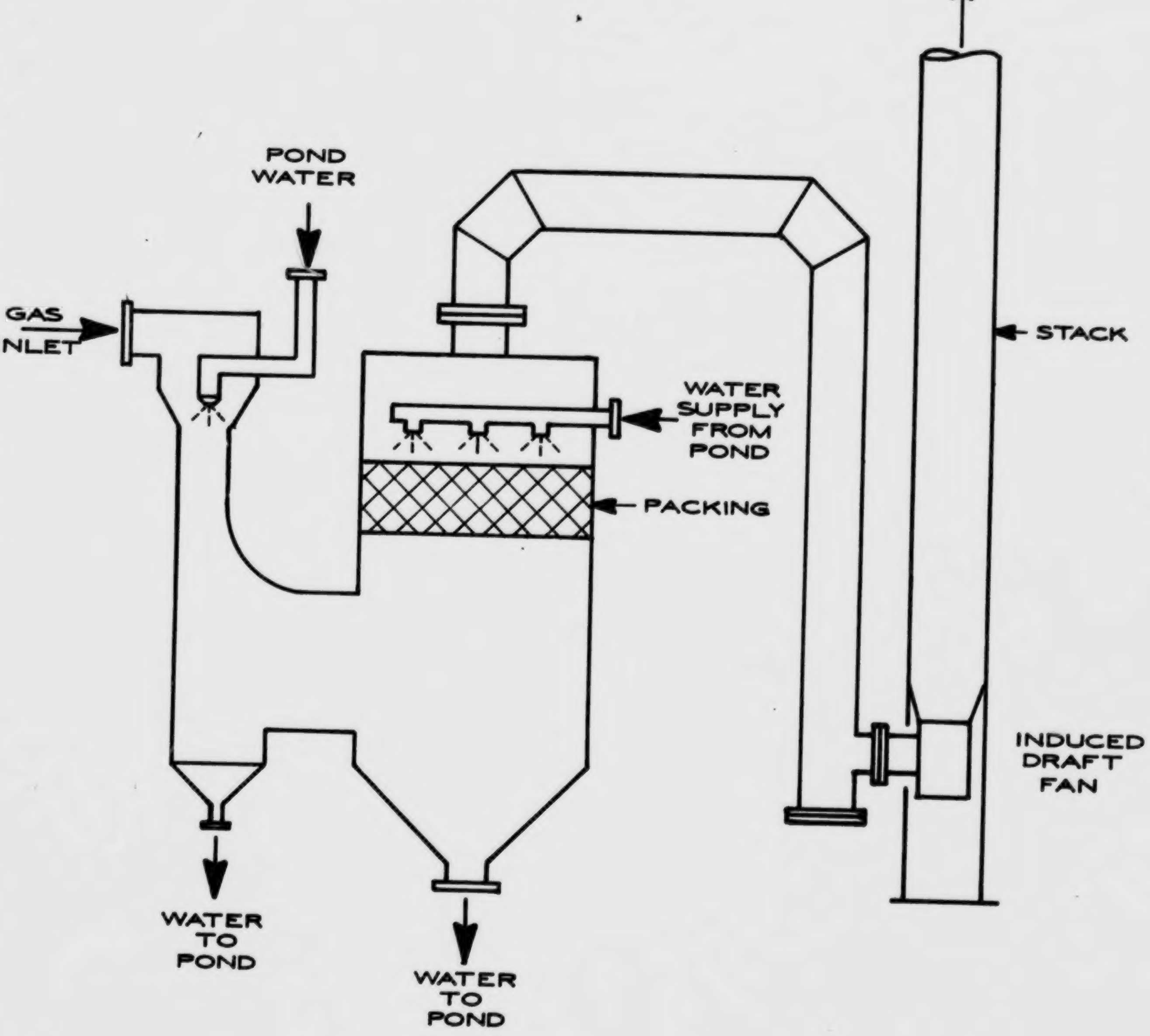

POND WATER
GAS INLET
WATER SUPPLY FROM POND
PACKING
STACK
INDUCED DRAFT FAN
WATER TO POND
WATER TO POND

TABLE 22

PROCESS DESCRIPTION FOR WET SCRUBBER FOR SPA PROCESS SPECIFICATION

The scrubber is to serve a 300 T/day Vacuum Evaporation superphosphoric acid plant. The scrubber is to treat all of the gases discharged from the process vents as follows:

(1) Barometric Condenser vents

(2) Hotwell vents

(3) Hot product cooling tank vent

The scrubber is to use gypsum pond water with the following characteristics:

	Design	*Min.*	*Max.*
Pond Water pH	2.0	1.2	2.2
Temperature, °F	80	55	88
SO_4*, wt.%*	0.15		
P_2O_5*, wt.%*	0.7		
H_2SiF_6*, wt.%*	0.63	0.25	1.0
Fluorine, wt.%	0.5	0.2	0.8

The scrubber is required to produce the specified performance when operating with water at "design" conditions. The scrubber manufacturer shall specify the circulation rate through the scrubber.

Materials of construction shall be limited to the following, which shall be selected for hydrofluoric acid service:

PVC

Rubber (below 160°F)

FRP (Dynel lined)

No metal parts shall be used where exposure to process gas or pond water may be significant.

Two scrubbers are to be considered as alternatives for this service.

A. Water-Induced Venturi

The water-induced Venturi scrubber shall be furnished so as to provide for a negative pressure of 1/4" w.c. at the inlet to the scrubber when processing gas at the design flow rate. The scrubber shall be equipped with an entrainment separator and a stub stack which extends to approximately 25 feet above grade.

The scrubber is to be located adjacent to the hotwells, to which it may be connected by approximately 20 ft. of suitable ductwork, which is to be furnished as a part of the scrubber installation. The other vents will be piped into the hotwell by the piping contractor.

B. Venturi with Optional Packed Section

The Venturi scrubber shall be furnished complete with fan, interconnecting ductwork and 25 ft. discharge stack. The fan shall be sized so as to provide sufficient static head for overcoming the pressure drop in the scrubber system and inlet ductwork.

The scrubber is to be located adjacent to the hotwells, to which it may be connected with 20 ft. of ductwork which is to be furnished as a part of this installation. In addition, an inlet damper and barometric damper shall be installed in the ductwork adjacent to the scrubber inlet for adjusting the gas flow through the scrubber.

TABLE 23

OPERATING CONDITIONS FOR WET SCRUBBER FOR SPA PROCESS COMBINED VENTS SPECIFICATION

Plant Capacity, ton/day	*300*
Acid Strength, Wt.% P_2O_5	
to Evaporator	*54*
from Evaporator	*72*
Fluorine Content, Wt.% F	
to Evaporator	*1.5*
from Evaporator	*0.4*

Scrubber inlet Streams *Source*	*Barometric Condenser*	*Hotwell*	*Cooling Chamber*	*Total to Scrubber*
Flow, ACFM	*287*	*2,120*	*264*	*2,671*
Temp., °F	*147.5*	*100*	*100*	*105*
Flow, SCFM	*250*	*2,000*	*250*	*2,500*
Moisture, Vol.%	*24*	*6.7*	*6.7*	*7.8*
Flow, DSCFM	*190*	*1,880*	*234*	*2,304*
Fluorine, ppm	*10*	*5*	*30*	*8.25*
Fluorine, lb/hr	*0.008*	*0.030*	*0.024*	*0.062*

Scrubber Outlet	
Flow, DSCFM	*2,304*
Temp., °F	*100*
Moisture, Vol. %	*6.7*
Flow, SCFM	*2,470*
Fluorine, ppm	*1.455*
Fluorine, lb/hr	*0.011*
Scrubber Efficiency, %	*82.5*
Estimated y', ppm	*1.4*
Estimated NTU requirement	*5.0*

TABLE 24

CAPITAL COST DATA (COSTS IN DOLLARS) FOR WATER-INDUCED VENTURI SCRUBBERS FOR SPA PROCESS PLANTS

	Nominal Efficiency			
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	1,335	2,671		
°F	105	105		
SCFM	1,250	2,500		
Moisture Content, Vol. %	7.8	7.8		
Effluent Contaminant Loading				
(ppm), Fluorine	8.25	8.25		
lb/hr , Fluorine	0.031	0.062		
Cleaned Gas Flow				
ACFM	1,310	2,620		
°F	100	100		
SCFM	1,235	2,470		
Moisture Content, Vol. %	6.7	6.7		
Effluent Contaminant Loading				
(ppm), Fluorine	1.455	1.455		
lb/hr, Fluorine	0.0055	0.011		
Cleaning Efficiency, %	82.5	82.5		
(1) Gas Cleaning Device Cost	6,383	8,603		
(2) Auxiliaries Cost				
(a) Fan(s)				
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost	2,467	2,673		
(a) Engineering				
(b) Foundations & Support	700*	760*		
(c) Ductwork	250*	325*		
(d) Stack				
(e) Electrical	250*	250*		
(f) Piping	300*	475*		
(g) Insulation				
(h) Painting				
(i) Supervision	300*	300*		
(j) Startup				
(k) Performance Test	375*	375*		
(l) Other				
(4) Total Cost	9,183	11,843		

*Items of installed cost, average of two bidders. Third bidder did not itemize.

FIGURE 20
CAPITAL COST FOR WATER INDUCED VENTURI SCRUBBERS
FOR SPA PROCESS PLANTS

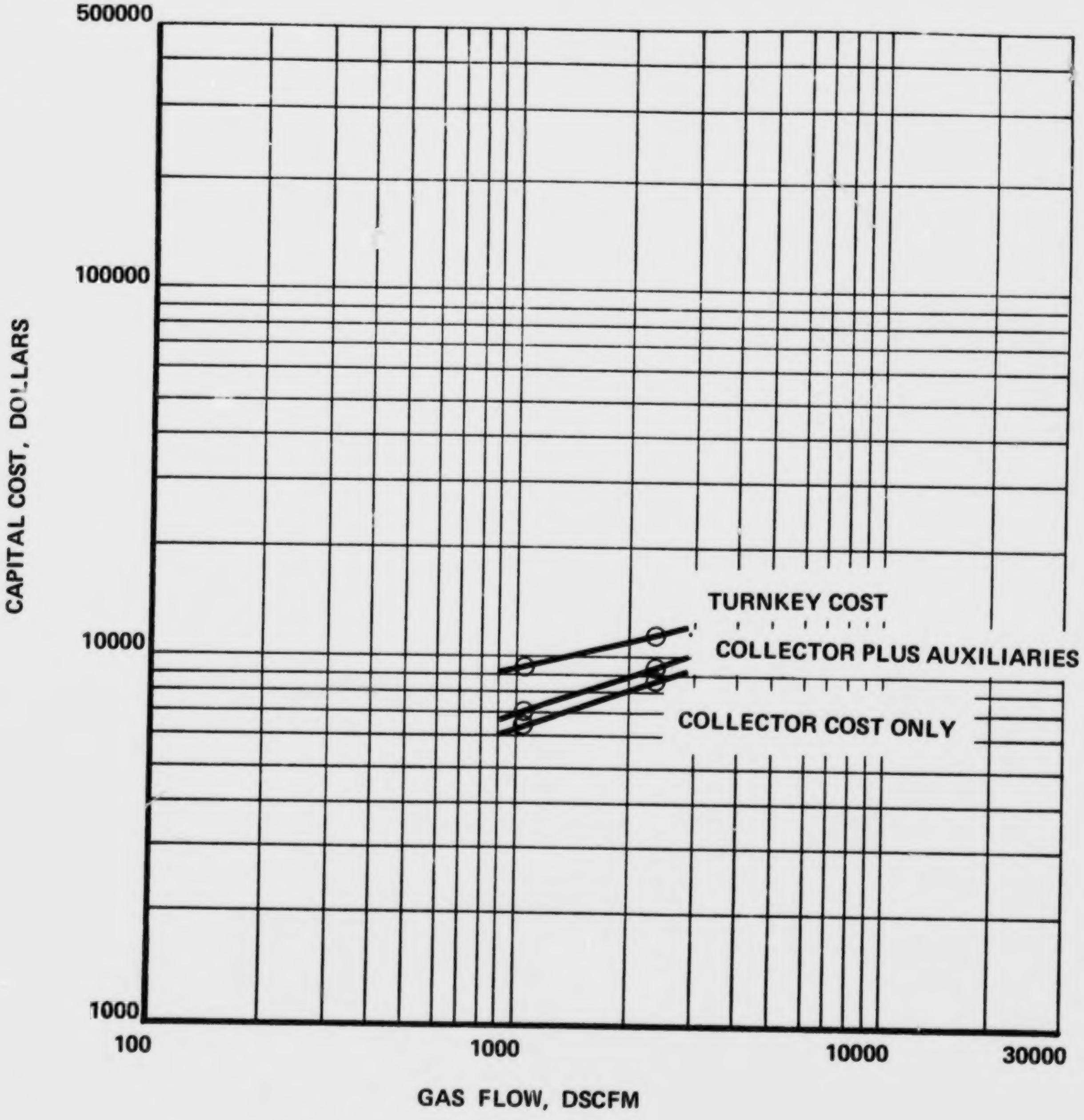


TABLE 25

CONFIDENCE LIMITS FOR CAPITAL COST OF WATER INDUCED VENTURI SCRUBBERS FOR SPA PROCESS PLANTS

Population – 5 Sample Size – 3

Capital Cost = $6,383

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$5,162	$7,602
75	4,018	8,747
90	2,258	10,507
95	630	12,135

Capital Cost = $8,603

Conf. Level, %	Lower Limit	Upper Limit
50	$6,940	$10,265
75	5,381	11,824
90	2,982	14,223
95	764	16,442

FIGURE 21

CONFIDENCE LIMITS FOR CAPITAL COST OF WATER INDUCED VENTURI SCRUBBERS FOR SPA PROCESS PLANTS

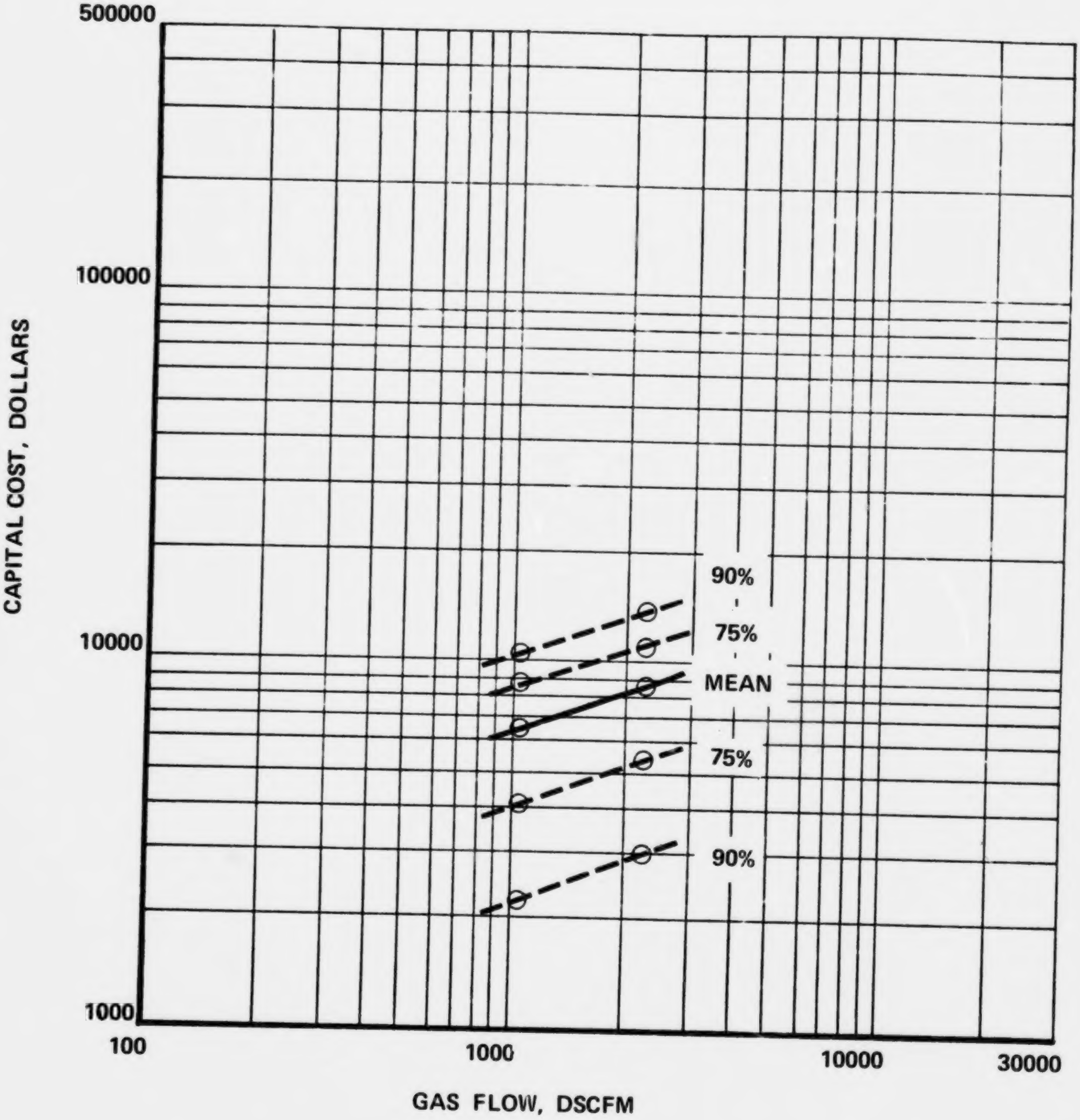


TABLE 26

CONFIDENCE LIMITS FOR INSTALLED COST OF WATER INDUCED VENTURI SCRUBBERS FOR SPA PROCESS PLANTS

Population – 5 Sample Size – 3

Installed Cost = $9,182

	Installed Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$ 7,999	$10,366
75	6,889	11,475
90	5,181	13,183
95	3,603	14,762

Installed Cost = $11,843

Conf. Level, %	Lower Limit	Upper Limit
50	$10,416	$13,270
75	9,078	14,608
90	7,019	16,667
95	5,115	18,570

FIGURE 22
CONFIDENCE LIMITS FOR INSTALLED COST OF WATER INDUCED VENTURI SCRUBBERS FOR SPA PROCESS PLANTS

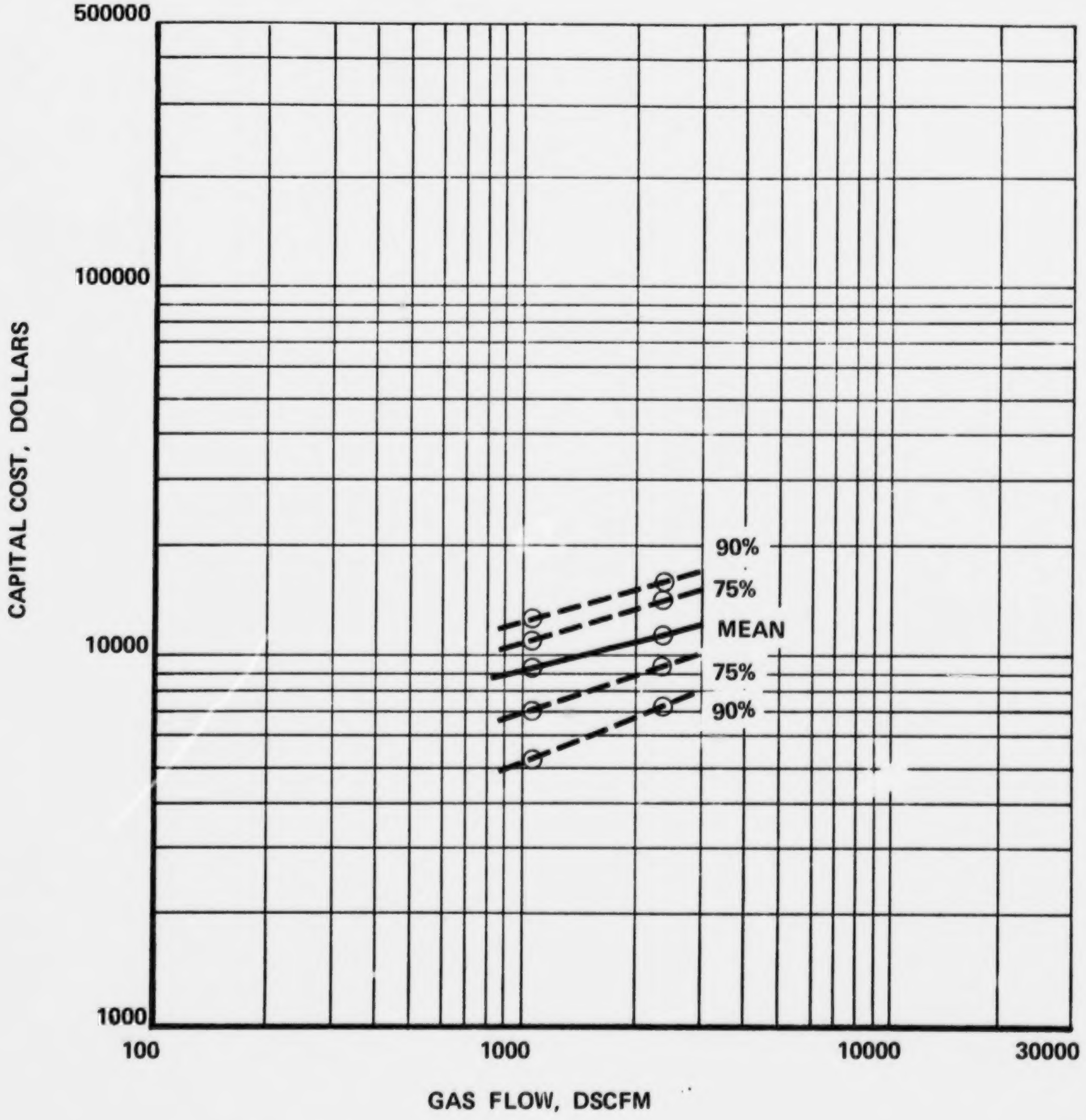


TABLE 27

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR WATER-INDUCED VENTURI SCRUBBERS FOR SPA PROCESS PLANTS

Operating Cost Item	Unit Cost	Nominal Efficiency			
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor		$ 200	$ 200		
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		74	76		
Replacement Parts					
Total Replacement Parts		442	540		
Utilities					
Electric Power	.011/kw-hr	$ 877	$1,425		
Fuel					
Water (Process)	.25/M gal				
Water (Cooling)	.05/M gal				
Chemicals, Specify					
Total Utilities		877	1,425		
Total Direct Cost		1,593	2,241		
Annualized Capital Charges		918	1,184		
Total Annual Cost		$2,511	$3,425		

FIGURE 23

ANNUAL COST FOR WATER INDUCED VENTURI SCRUBBERS FOR SPA PROCESS PLANTS

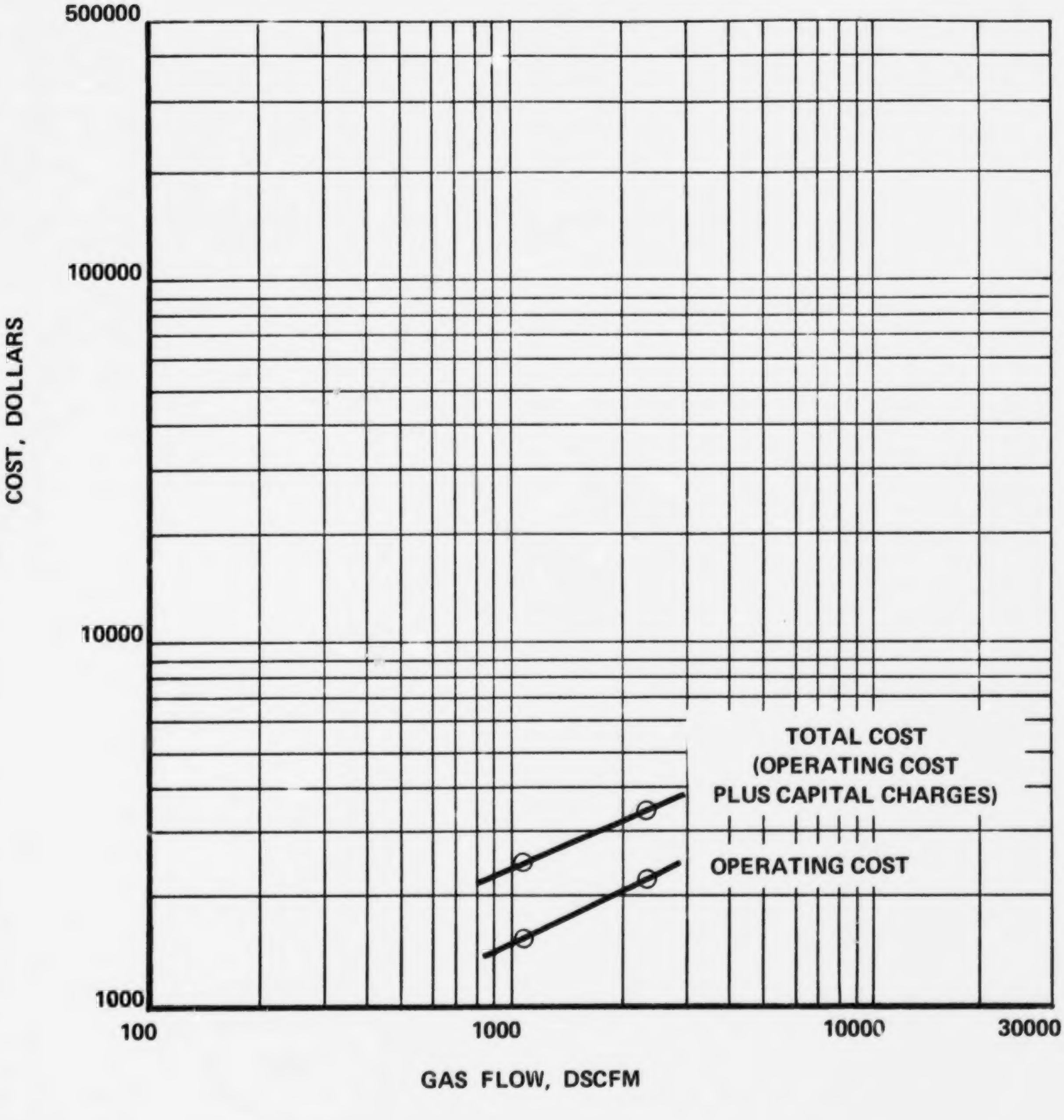


TABLE 28

CAPITAL COST DATA (COSTS IN DOLLARS) FOR VENTURI SCRUBBER WITH PACKED SECTION FOR SPA PROCESS PLANTS

	Nominal Efficiency			
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	1,335	2,671		
°F	105	105		
SCFM	1,250	2,500		
Moisture Content, Vol. %	7.8	7.8		
Effluent Contaminant Loading				
(ppm), Fluorine	8.25	8.25		
lb/hr, Fluorine	0.031	0.062		
Cleaned Gas Flow				
ACFM	1,310	2,620		
°F	100	100		
SCFM	1,235	2,470		
Moisture Content, Vol. %	6.7	6.7		
Effluent Contaminant Loading				
(ppm), Fluorine	1.455	1.455		
lb/hr , Fluorine	.0055	.011		
Cleaning Efficiency, %	82.5	82.5		
(1) Gas Cleaning Device Cost	6,292	7,742		
(2) Auxiliaries Cost	4,167	4,683		
(a) Fan(s)	3,100*	3,400*		
(b) Pump(s)	800*	800*		
(c) Damper(s)	500*	575*		
(d) Conditioning, Equipment	--	--		
(e) Dust Disposal Equipment	--	--		
(3) Installation Cost	13,597	16,155		
(a) Engineering	2,600*	2,650*		
(b) Foundations & Support	2,750*	3,500*		
(c) Ductwork		--		
(d) Stack		--		
(e) Electrical	1,500*	1,750*		
(f) Piping	1,350*	1,550*		
(g) Insulation				
(h) Painting				
(i) Supervision				
(j) Startup				
(k) Performance Test	2,150*	2,150*		
(l) Other				
(4) Total Cost	24,056	28,580		

*Items of installed cost, average of two bidders. Third bidder did not itemize.

FIGURE 24

CAPITAL COST FOR VENTURI SCRUBBERS WITH PACKED SECTION FOR SPA PROCESS PLANTS

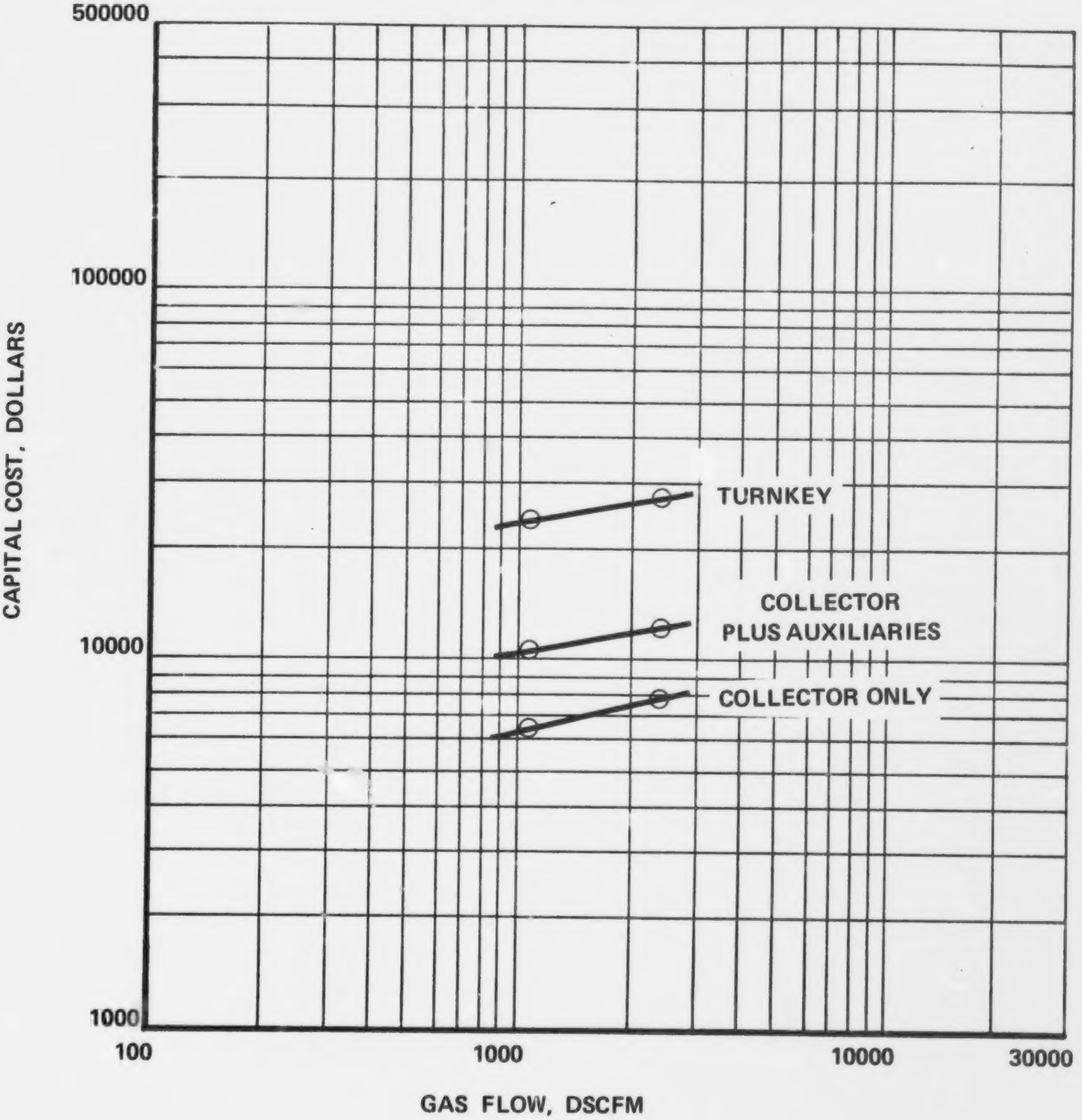


TABLE 29

CONFIDENCE LIMITS FOR CAPITAL COST OF VENTURI SCRUBBERS WITH PACKED SECTION FOR SPA PROCESS PLANTS

Population Size – 5 Sample Size – 3

Capital Cost = $6,292

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$5,515	$7,069
75	4,786	7,797
90	3,665	8,918
95	2,629	9,954

Capital Cost = $7,742

Conf. Level, %	Lower Limit	Upper Limit
50	$6,713	$8,770
75	5,748	9,734
90	4,264	11,218
95	2,892	12,591

FIGURE 25

CONFIDENCE LIMITS FOR CAPITAL COST OF VENTURI SCRUBBERS WITH PACKED SECTION FOR SPA PROCESS PLANTS

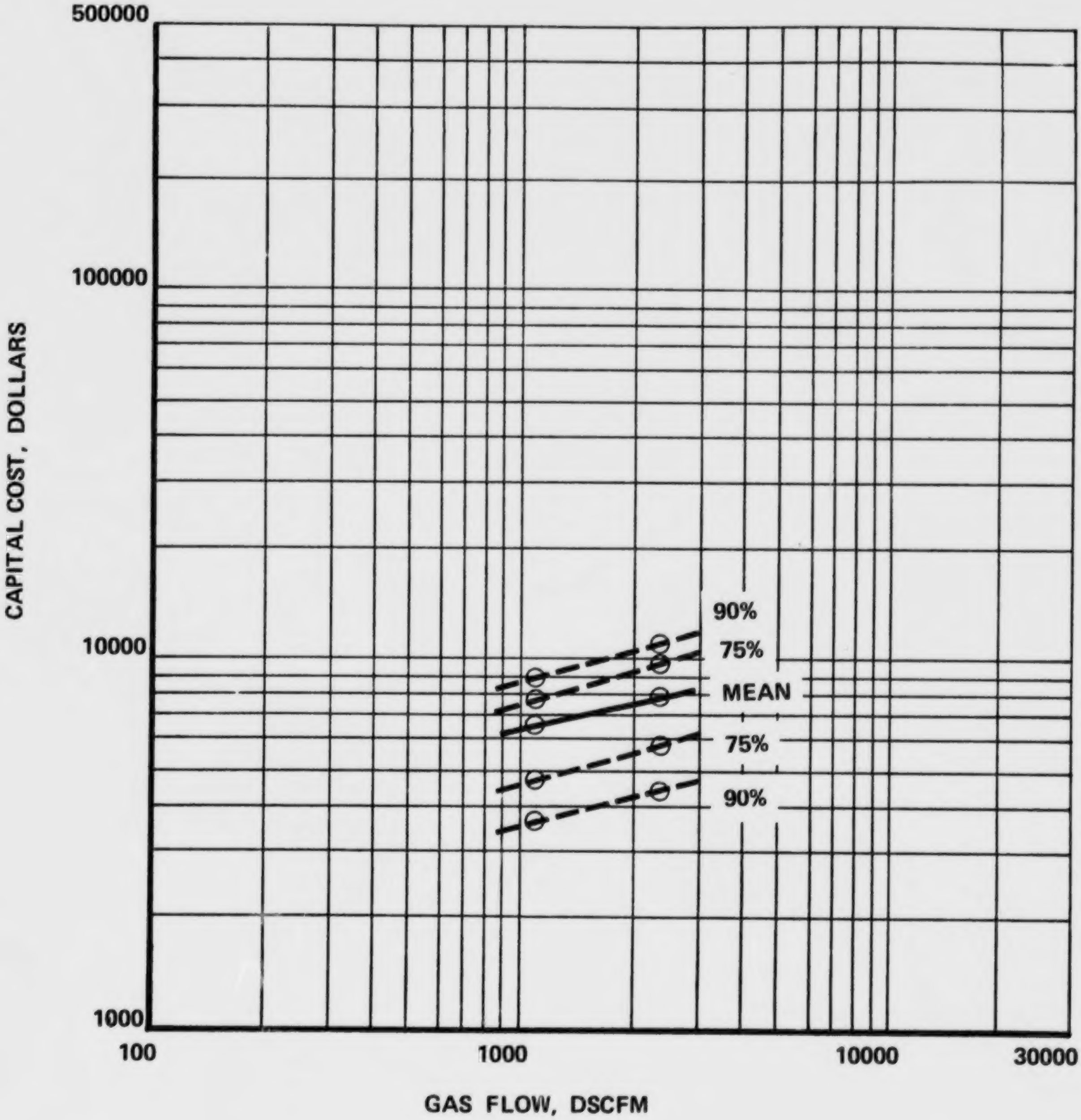


TABLE 30

CONFIDENCE LIMITS FOR INSTALLED COST OF VENTURI SCRUBBERS WITH PACKED SECTION FOR SPA PROCESS PLANTS

Population Size – 5 Sample Size – 3

Installed Cost = $24,055

	Installed Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$22,423	$25,687
75	20,893	27,217
90	18,538	29,572
95	16,361	31,749

Installed Cost = $28,580

Conf. Level, %	Lower Limit	Upper Limit
50	$26,148	$31,012
75	23,868	33,292
90	20,359	36,801
95	17,114	40,046

FIGURE 26

CONFIDENCE LIMITS FOR INSTALLED COST OF VENTURI SCRUBBER WITH PACKED SECTION FOR SPA PROCESS PLANTS

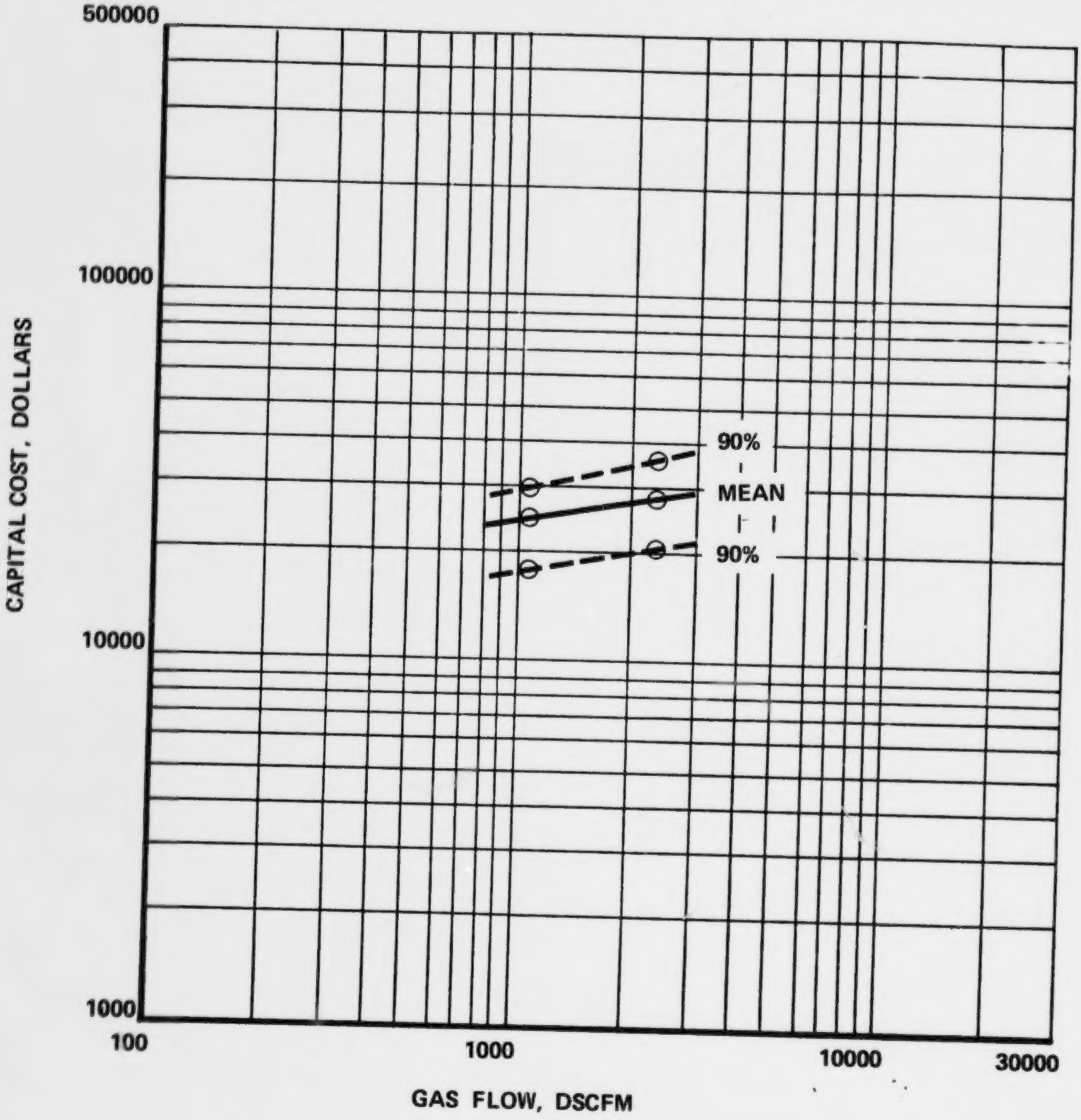


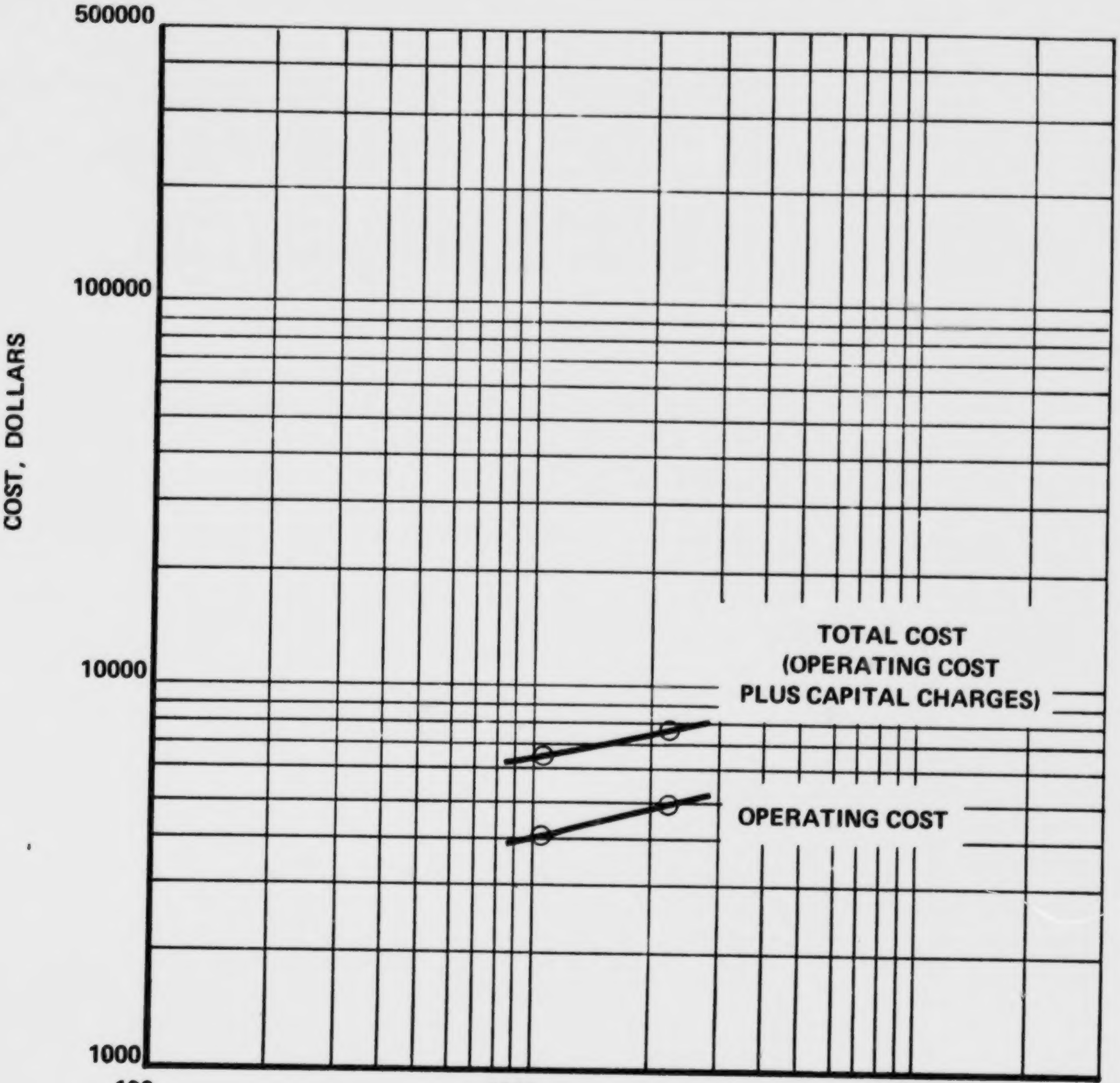
TABLE 31

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR VENTURI SCRUBBER WITH PACKED SECTION FOR SPA PROCESS PLANTS

Operating Cost Item	Unit Cost	Nominal Efficiency		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor		$1,227	$1,227		
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		700	608		
Replacement Parts					
Total Replacement Parts		612	728		
Utilities					
Electric Power	$.011/kw-hr	610	1,164		
Fuel					
Recirculating Pond Water	$.05/M gal	377	604		
Chemicals, Specify					
Total Utilities		987	1,768		
Total Direct Cost		3,526	4,331		
Annualized Capital Charges		2,406	2,858		
Total Annual Cost		5,932	7,189		

FIGURE 27

ANNUAL COST FOR VENTURI SCRUBBERS WITH PACKED SECTION FOR SPA PROCESS PLANTS



TRIPLESUPERPHOSPHATE GRANULATION PLANTS

Triplesuperphosphate is manufactured by *acidulation* of phosphate rock with phosphoric acid. The two processes in common use differ in the method of solidifying and drying the solid product. The older method discharges the precipitated material onto a slow-moving belt or *den* and produces a solid aggregate of varying particle size. This material is cured for thirty days or more while the reactions go to completion, and is sold as run-of-pile (abbreviated ROP), triplesuperphosphate.

The den method of production is being replaced by direct granulation of the product as it is formed. This discussion deals exclusively with the granulation process (GTSP) as it is applied to Florida phosphate operations.

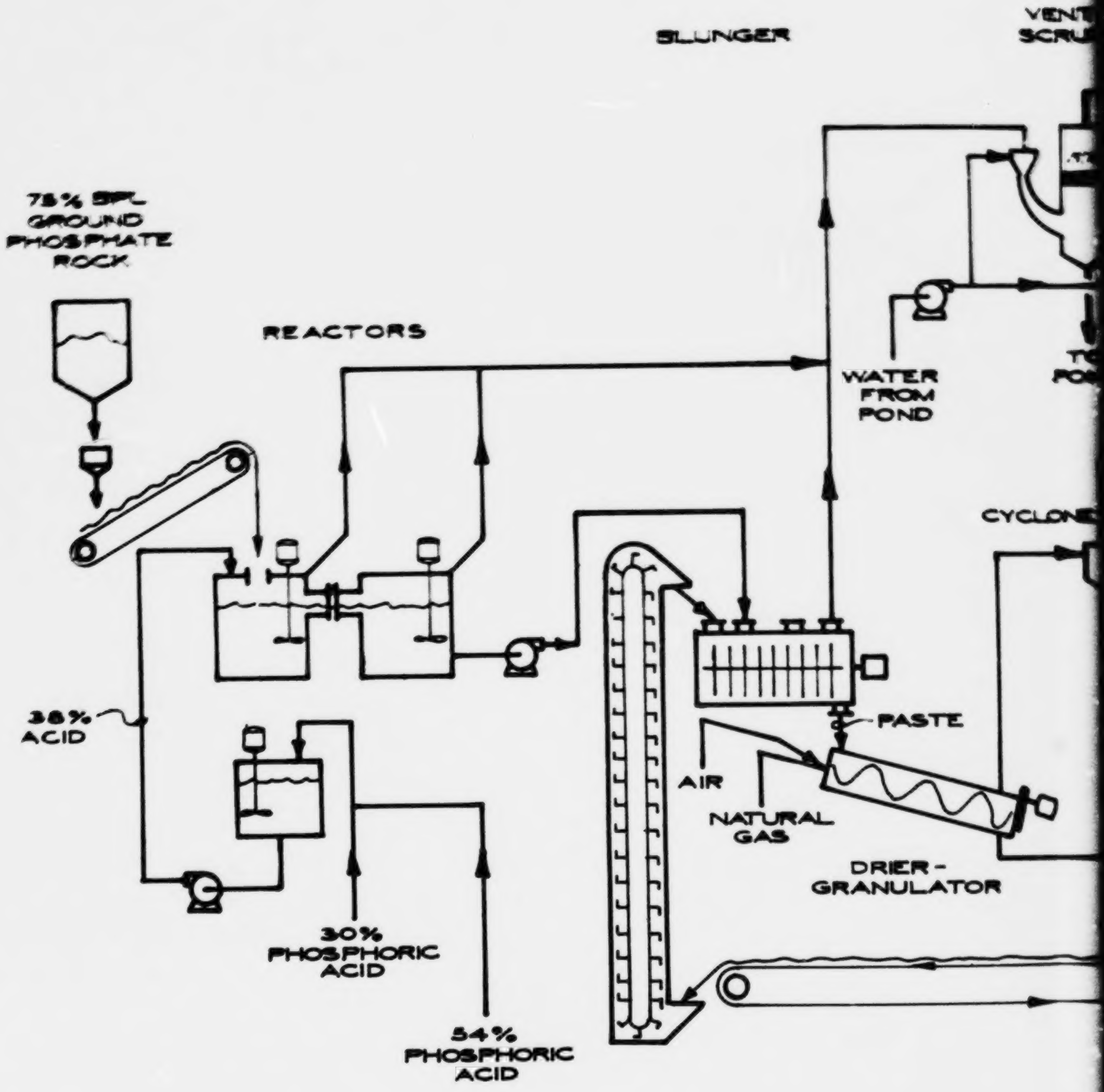
PROCESS DESCRIPTION

Figure 28 is a schematic flow diagram for a typical triplesuperphosphate granulation plant.(16)

Finely ground commercial rock is charged to the process from a ball or roller milling and screening operation. The rock is metered onto a belt conveyer at a fixed weight rate of flow and discharged into a reactor vessel. Thirty-eight percent P_2O_5 phosphoric acid is charged into the same reactor and formed into a slurry with the phosphate rock. Ordinarily, this intermediate strength acid is made by blending concentrated acid from the vacuum evaporation section of a phosphoric plant at 54 wt. % P_2O_5 strength with 30 wt. % phosphoric acid taken directly from the digester section of the phosphoric acid plant. These two streams are premixed in a combination mixer and hold tank which provides some surge capacity.

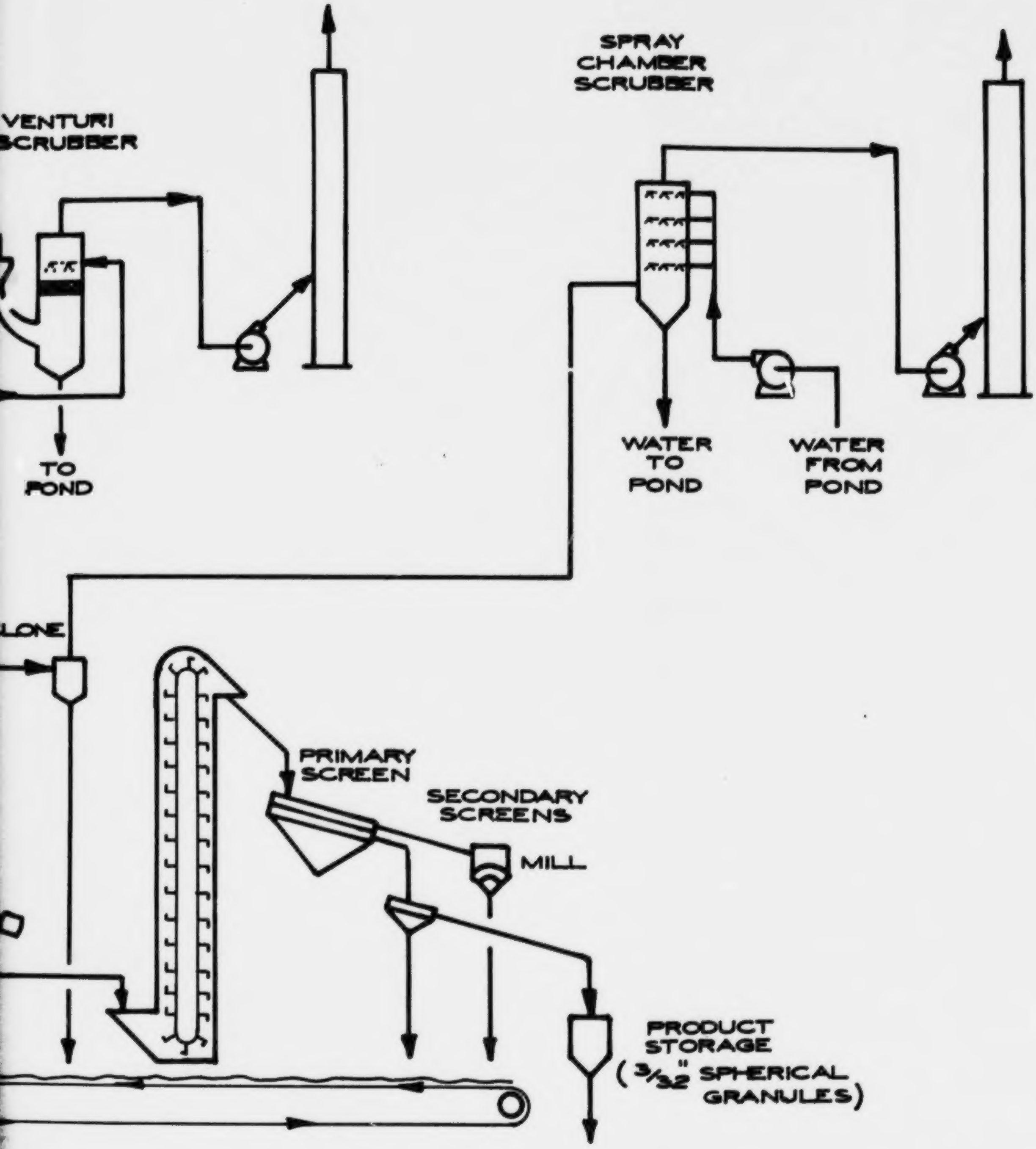
The slurry is held and agitated until the reaction is near completion, and the mixture overflows into a second reaction chamber where the acidulation reaction is completed. Both reactor vessels are vented through a scrubber or combination of scrubbers to the atmosphere. The slurry leaving the reactors contains all of the product material plus additional free water which will be vaporized in the drier. The liquid reactor effluent is fed into a horizontal mixing vessel called a *blunger* where it is admixed with a recycled solid material of smaller size than the desired product granules.

The blunger serves to further pulverize the undersized solids and mix it into a paste with the liquid reactor product. This paste is discharged into a


SLUNGER
VENT
SCRU
75% BPL
GROUND
PHOSPHATE
ROCK
REACTORS
WATER
FROM
POND
TO
PO
CYCLON
PASTE
AIR
NATURAL
GAS
DRIER-
GRANULATOR
38%
ACID
30%
PHOSPHORIC
ACID
54%
PHOSPHORIC
ACID

FIGURE

FLOW DIAGRA
TRIPLE SUPERP
GRANULATIO


VENTURI
SCRUBBER
TO
POND
SPRAY
CHAMBER
SCRUBBER
WATER
TO
POND
WATER
FROM
POND
LONE
PRIMARY
SCREEN
SECONDARY
SCREENS
MILL
PRODUCT
STORAGE
(3/32" SPHERICAL
GRANULES)

URE 28

AGRAM FOR
PERPHOSPHATE
ATION PLANT

granulator-dryer from which water vapor, products of combustion and granular solid products are discharged. The blunger and reaction tanks are usually ventilated through a common scrubber and the gas is discharged to the atmosphere. The dryer is heated by a direct-fired gas or oil burner. The solid product leaving the dryer is discharged into a bucket elevator and lifted to a primary screening device. The coarse material, larger than the desired product granule size, is discharged from the screen into a mill which pulverizes it and drops it onto a conveyer for return to the blunger. An intermediate screen retains materials of approximately the desired product size range and drops the fines through onto the same conveyer belt for return to the blunger. The overflow from the product screen is discharged to a secondary screen where a final separation is made between the product material and fines. Again, the fines are returned to the blunger by the common conveyer belt. The granular product is transported to a storage building where it is retained for six to eight days while the "curing" reactions are completed.

The combustion gas from the granulator-dryer is discharged into a cyclone where the fine TSP particles entrained are separated and returned to the blunger by way of a common fines conveyer belt. The gas discharge from the cyclone is vented to the atmosphere through a scrubbing system which serves to remove particulate matter, chlorides, and in some cases sulfur dioxide.

CHEMISTRY OF THE PROCESS

The basic chemistry is represented in a superficial way by equation (2) as follows:

$$\mathrm{Ca_3\ (PO_4)_2 + 4H_3PO_4 + 3H_2O \rightarrow 3\,CaH_4\ (PO_4)_2 \cdot H_2O} \qquad (2)$$

Phosphate Rock	Phosphoric Acid	Water	Monocalcium Phosphate

This equation is oversimplified both by the omission of the fluorine component of the fluorapitite rock, and also by the omission of the silica diluent. A better representation is given in equation (22):

$$\mathrm{3\ [Ca_3(PO_4)_2]\ CaF_2 + 14H_3PO_4 + HQH_2O \rightarrow 10CaH_4\ (PO_4)_2 \cdot H_2O + 2HF} \qquad (22)$$

Fluorapatite + Phosphoric Acid + Water →Monocalcium Phosphate → Hydrofluoric Acid

Preceding page blank

where the fluorine evolution as HF is indicated. However, as in the case of the wet phosphoric acid manufacture, the chemical equilibrium between HF and silicon tetrafluoride is represented by equations (23) and (24):

$$2HF + SiO_2 \rightleftarrows H_2O + SiF_4 \quad (23)$$

Hydrofluoric Acid + Silica ⇄ Water + Silicon Tetrafluoride

and

$$SiF_4 + 2HF \rightleftarrows H_2SiF_6 \quad (24)$$

Silicon Tetrafluoride + Hydrofluoric Acid ⇄ Fluorosalicic Acid

favors the emission of SiF_4 at low temperatures.

A number of side-reactions take place and produce by-products such as

$$Ca_3(PO_4)_2 + H_3PO_4 \xrightarrow{+ H_2O} 3CaHPO_4 \cdot 2H_2O$$

Tricalcium Phosphate + Phosphoric Acid Water → Calcium Metaphosphate (25)

None of these side reactions are of significance in the consideration of air pollution emissions, however.

RAW MATERIALS

The principal raw material for the process is ground-to-size phosphate rock similar to that used for wet process phosphoric acid manufacture. Figure 2 indicates the ore benefaction and milling and sizing steps which precede the GTSP process.

Figure 29 shows a flow scheme of a phosphate rock milling operation. Granular rock at about 1/4 inch size is delivered by rail or truck from a benefaction plant to a loading hopper and loaded into a storage silo. Rock is charged from the silo to a ball mill in which it is ground to 80% through a 100 mesh screen or finer.(1)

The mill discharge is conveyed pneumatically to a centrifugal classifier which separates the ground rock according to particle size. The largest fraction,

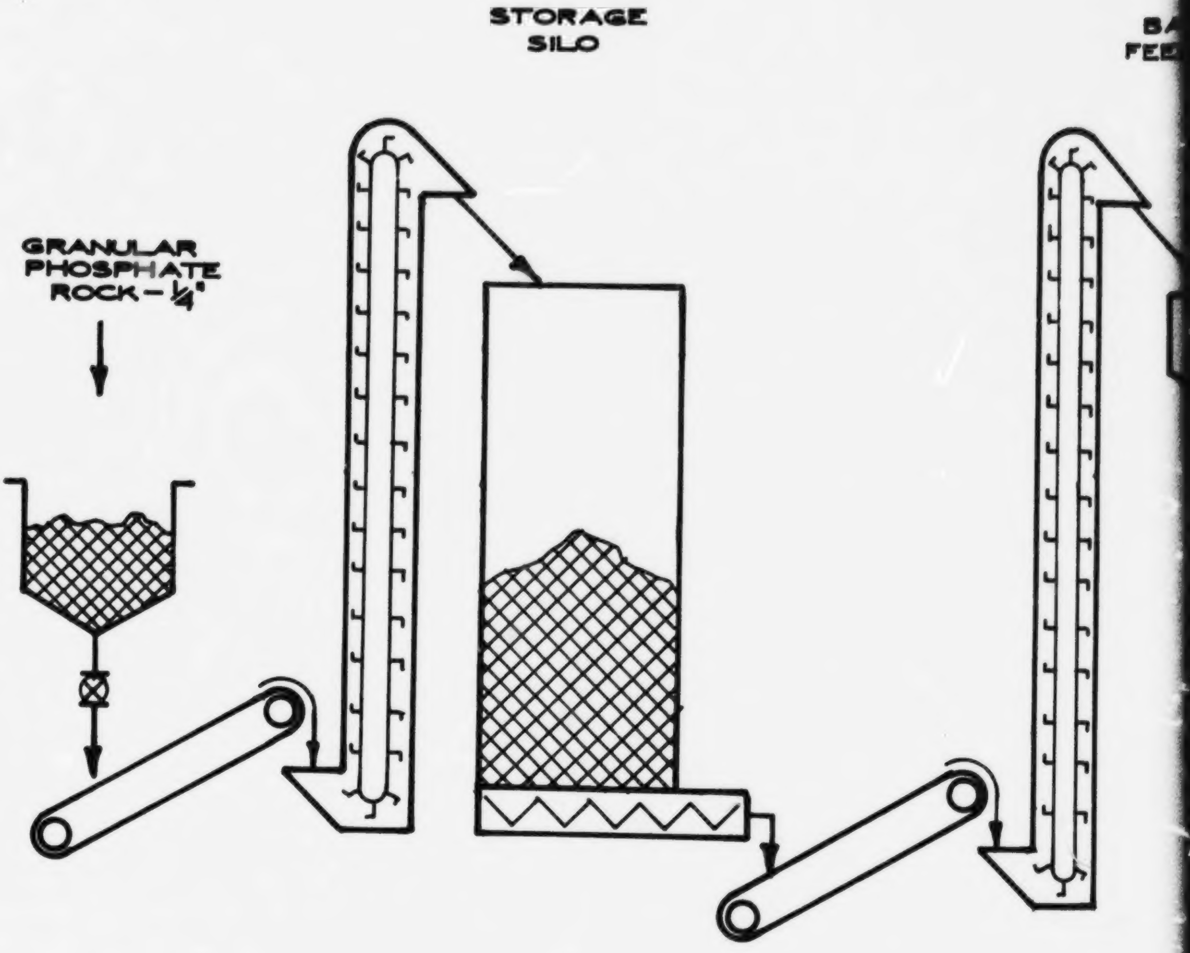
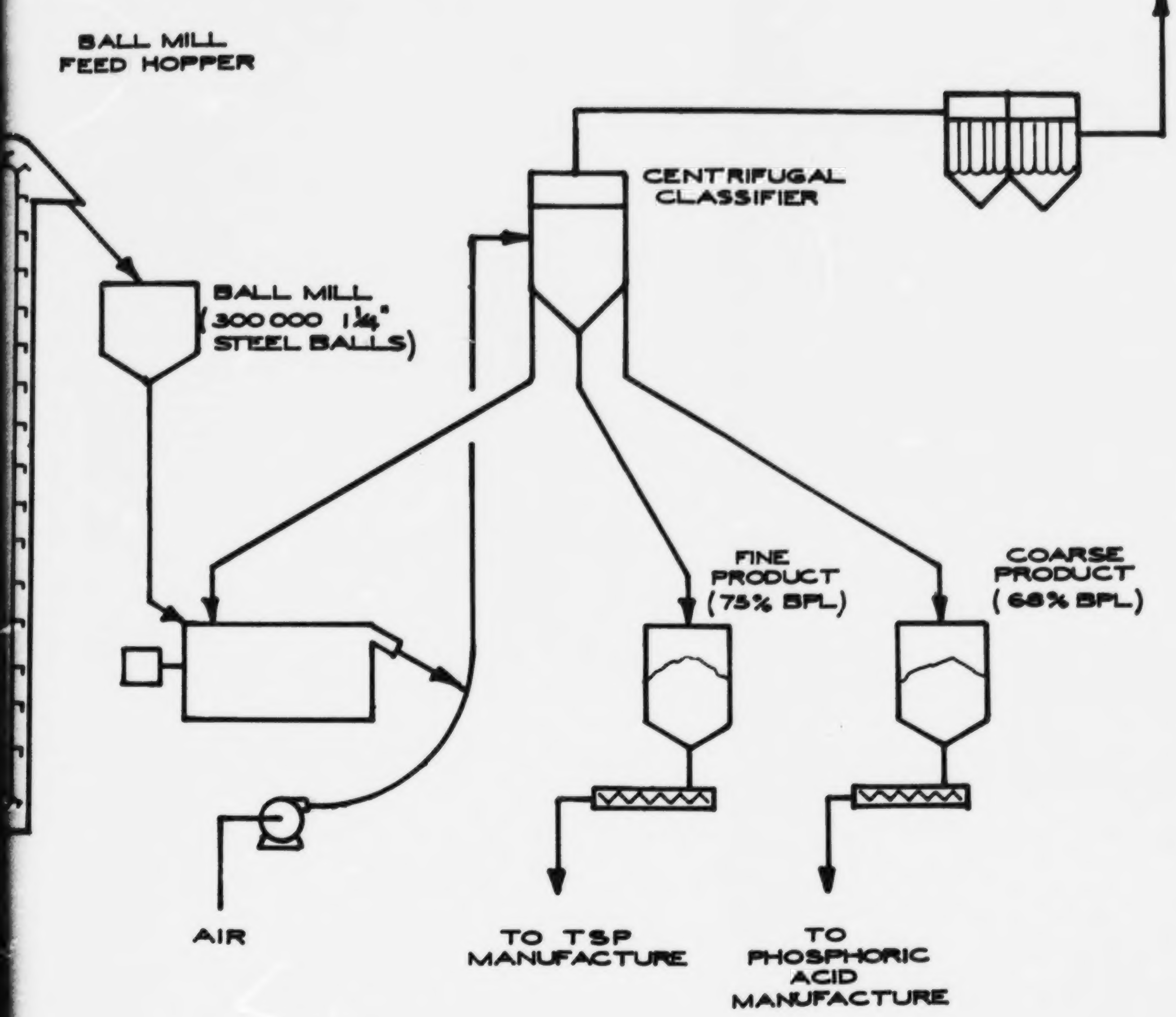

STORAGE SILO
BA
FEE
GRANULAR PHOSPHATE ROCK - ¼"

FIGURE 2

FLOW SCHEME OF
PHOSPHATE ROCK MILL


BALL MILL
FEED HOPPER
CENTRIFUGAL
CLASSIFIER
BALL MILL
(300 000 1¼"
STEEL BALLS)
FINE
PRODUCT
(75% BPL)
COARSE
PRODUCT
(68% BPL)
AIR
TO TSP
MANUFACTURE
TO
PHOSPHORIC
ACID
MANUFACTURE

IGURE 29

EME OF 6000 T/D
CK MILLING OPERATION

TABLE 32

MATERIAL BALANCE OF GTSP PROCESS FOR 250 TON/DAY

	Acid Feed		Rock Feed		Product GTSP	
	Ton/day	lb/hr	Ton/day	lb/hr	Ton/day	lb/hr
P_2O_5	175.0	14,600	75.0	6,250	250.0	20,850
CaO	–	–	109.8	9,150	109.8	9,150
F	4.3	416	8.3	685	9.1*	800*
SO_3	5.6	460	2.6	212	8.1	672
Al_2O_3	6.5	540	1.9	163	8.4	703
Fe_2O_3	2.6	218	2.4	205	5.0	422
Other	1.3	108	25.2	2,100	13.7**	1,225**
Water	264.7	22,058	2.3	190	109.9***	9,115***
Total	460.0	38,400	227.5	18,955	515.0	42,937

*Based on retention of 72.5 of F in solid product after storage.

**Based on evolution of CO_2 and water as from decomposition of carbonates and organic matter.

***Product moisture taken at 1 wt.% free H_2O.

Preceding page blank

greater than 100 mesh, is returned to the mill.

Each of the ground products is transferred to intermediate storage prior to charging to the next process in line.

Phosphoric acid makes up the only other raw material used. This is obtained by blending 30% acid and 54% acid, which produces a total acid with somewhat more fluorine in it than would be the case were all 54% acid used and blended with water. However, no serious error will be introduced in this discussion if the composition of 54% acid given in Table 7 is used.

Table 32 gives a material balance for a hypothetical process with a production rate of 250 Ton/day P_2O_5. For purposes of preparing this table, it was assumed that there was no loss of phosphorus values (i.e. that particulate losses were substantially all returned to the process).

Several notations in Table 32 are significant with regard to air pollution emissions. The following approximate fluorine balances were reported values by Teller(6) for the ROP process and by Chemico(17) for the GTSP process.

	ROP % of Total F	GTSP % of Total F
Fluorine retained in product	65	72.5
Fluorine discharged in reactor - dryer - cooler	33.5	27
Fluorine discharged in storage	1.5	0.5
	100.0	100.0

The "other" materials contained in the acid and rock feed consist of

Silicates
Carbonates
Organics

as reported in the WPPA section of this report.

As the acidulation process takes place in the reactor, it is likely that the carbonates and organics decompose and are discharged from the reactor and dryer vents as CO_2 and water vapor. In addition, some of the fluorine vaporizes as SiF_4 and carries some of the silica out of the system.

TABLE 33

MATERIALS EVOLVED AS GASES FROM TRIPLESUPERPHOSPHATE MANUFACTURE FOR 250 TON/DAY PROCESS

	Ton/day	lb/hr	SCFM
Water			
From Acid	157.1	13,133	4,600
From Organics	2.8	237	82
Total	159.9	13,370	4,682
CO_2			
From Carbonates	6.6	545	78
From Organics	13.7	1,143	164
Total	20.3	1,688	242
SiF_4			
Fluorine	3.4	283	–
Silicon	1.25	104	–
Total	4.65	387	23.5

Table 33 presents a calculated breakdown of the CO_2, water and fluorine balances for the reactor and dryer.

GASEOUS DISCHARGES

Four principal emission points are common in GTSP plants:

1. The reactor vents
2. Blunger vents
3. Dryer gas effluent.
4. Dust exhaust

This study is concerned with assembly of cost information only for the dryer vents. However, some discussion of the other effluents is included for completeness.

REACTORS

A principal source of emissions in the GTSP process is the reactor vessels. This is ventilated at relatively high rates because of the large open areas required for the continuous addition of ground rock. The rate of ventilation depends largely on the mechanical design of the reactor, in that there is little emission of gas from the reaction.

For purposes of obtaining cost data, the following gas flows were taken as typical of current design practice for the reactor-granulator combination:

Ton/day P_2O_5	Gas Flow, SCFM
250	15,000
500	20,000

Although the reaction of the rock and acid is strongly exothermic, and temperatures as high as 180° F are experienced in the reactor, the temperature of the ventilating air leaving the reactor seldom exceeds 140° F. The fluorine emission from the reactor is relatively high, and may reach 25 lb/ton P_2O_5.(6)

The gaseous contaminant is reported to be SiF_4, as the HF reacts with the excess of phosphate rock and is reabsorbed.[6]

For a 250 Ton/day process as illustrated in Tables 32 and 33, the reactor vent is assumed to contain about 100 mg/SCF fluorine as SiF_4, and 15 gr/SCF particulate rock dust.

The dryer vent is a second significant source of emissions in the GTSP process. The dryer receives granulated product from the granulator, along with a large loading of recycled fine material. Recycled solids/net product ratios on the order of 12/1 are common. This limits the capacity of the dryer for production of GTSP to a lower value than for DAP, where recycle/net solids ratios are on the order of 5/1.

The dryer is usually operated with counter-current contact between the flue gas and wet solids. The dryer may be gas or oil fired, and the products of combustion are tempered by dilution with ambient air to reduce the temperature below 1000° F before contact with the product. The exhaust gases from the dryer are again diluted somewhat with ambient air at the dryer outlet in order to reduce the moisture content and temperature.

Gas flows for GTSP dryers are likely to approximate those for DAP dryers handling the same moisture content. In fact, the same process equipment is often used interchangeably for DAP and GTSP manufacture. In this study, the gas flows were chosen so as to correspond exactly with the flowing volumes for DAP processes in order to obtain directly comparable prices.

POLLUTION CONTROL CONSIDERATIONS

Much of the discussion of pollution control devices for WPPA processes is also appropriate to the GTSP process. Gas absorption is of prime concern (the limitations on fluorine emission under present Florida law are 0.15 lb F/ton P_2O_5 for all process vents, and 0.05 lb F/ton for auxiliary equipment and storage, as given in Table 9). This requires reduction of the fluorine content to levels approaching those in equilibrium with gypsum pond water, and also necessitates the use of scrubbing equipment which is tolerant to deposits of gelatinous SiO_2 formed by reaction of the SiF_4 with pond water.

The scrubber specifications have been set on the basis that the dryer emission can produce one-half the total fluorine emission allowed under the Florida law, or 0.075 lb/ton P_2O_5. This amounts to 0.075 X 250 = 18.75

lb/day for the smaller of the two plants. The particulate matter entering the scrubber is presumed to be 3.5 wt. % fluorine and this has been taken into account in allowing a total weight of 18 lb/day F. This case is described as a "low efficiency" case with respect to fluorine removal. A second higher efficiency is defined by requiring the fluoride emission reduction to a level half as high.

Particulate emissions from the dryer are high, and an efficient particulate scrubber is required to reduce the particulates to a level comparable to the process weight limitations for particulates in general, or, to a level near the visible threshold. A concentration of 0.01 gr/SCF was taken as an approximation to the visible threshold.

Both of the particulate specifications require the device chosen to be capable of collecting particulate matter at high efficiency, and doing so without plugging with solids. These circumstances suggest that a Venturi-cyclonic of conventional design should be chosen as the primary collecting device. The low allowable emission of fluorine suggests that the Venturi-cyclonic alone may not be suitable for the complete absorption job.

Cross-flow or counter current-flow packed scrubbers, mobile packing scrubbers, or other gas absorption devices might be required as "tail gas" scrubbers following the Venturi to obtain the higher absorption efficiencies required.

TABLE 34

PROCESS DESCRIPTION FOR GRANULAR TRIPLESUPERPHOSPHATE SPECIFICATION

This specification describes a Venturi-packed scrubber for particulate removal and fluorine absorption service on a GTSP dryer. The particulate removal function is to be carried out in a Venturi section of conventional design using gypsum pond water as the scrubbing medium. Removal of entrained water and absorption of fluorides is to be accomplished in a second scrubbing stage containing a fixed plastic packing material. In addition to these functions, the scrubber must circulate sufficient pond water to reduce the temperature of the effluent gas to approximately 100°F in order to reduce the limiting vapor pressure of fluoride-containing gases.

Two efficiency levels are specified with respect to fluoride removal. These are to be met with pond water at the "design" condition specified below:

	Design	*Min.*	*Max.*
Pond water ph	*2.0*	*1.2*	*2.2*
Temperature, °F	*80**	*55*	*88*
SO_4*, wt.%*	*0.15*	–	–
P_2O_5*, wt.%*	*0.1*	–	–
H_2SiF_6*, wt.%*	*0.63*	*0.25*	*1.0*
Fluorine, wt.%	*0.5*	*0.2*	*0.8*

The water flow shall be calculated by the supplier to provide the proper heat balance.

Materials of construction shall be limited to the following, which shall be specified for hydrofluoric acid service:

PVC

Rubber-lined steel

FRP (Dynel lined)

No metal parts shall be used where exposure to process gas or pond water is likely to be significant.

**NOTE: Pond water average temperatures as high as 90°F are common and would require a more effective scrubber than specified here.*

Mechanical

The scrubber shall be furnished complete with interconnecting ductwork (if any) between the Venturi section and the packed section. The scrubber shall be equipped with a pressure regulating device, such as a barometric damper or variable throat mechanism which shall operate to minimize pressure fluctuations at the dryer.

A fan capable of overcoming the pressure drop across the scrubber stages and all ductwork shall be supplied as an auxiliary, and shall be located adjacent to the scrubber. A separate hotwell, enclosed to minimize fluorine loss, a return pump, 50 foot high stack and ductwork comprise the system.

The scrubber and all auxiliaries are to be located outside, adjacent to the dryer.

TABLE 35

OPERATING CONDITIONS FOR GRANULAR TRIPLESUPERPHOSPHATE WET SCRUBBER SPECIFICATION

Plant Capacity, ton/day P_2O_5	250	400
Plant Capacity, ton/day TSP	515	825
Process Weight, ton/hr	40	65
Gas to Scrubber		
Flow, DSCFM(1)	30,000	48,000
Flow, SCFM(1)	33,800	54,000
Flow, ACFM(1)	40,900	65,300
Temp., °F	180	180
Moisture, vol.%	12.64	12.64
Particulate, lb/hr	116	185
Particulate, gr/SCF	0.4	0.4
Fluorine F⁻, ppm	195	195
Fluorine, lb/hr	25	40
Gas from Scrubber		
Flow, DSCFM	30,000	48,000
Flow, SCFM	32,000	51,200
Flow, ACFM	34,000	54,000

	Mod. Eff.	High Eff.	Mod. Eff.	High Eff.
Temp., °F	100	100	100	100
Moisture, vol.%	6.47	6.47	6.47	6.47
Particulate, gr/SCF	0.02	0.02	0.02	0.02
Particulate, lb/hr	5.4	5.4	8.8	8.8
Fluorine F⁻, ppm	5.2	2.6	5.2	2.6
Fluorine, lb/hr	0.5(2)	0.25(3)	0.8(2)	0.4(3)
Particulate efficiency, %	95	95	95	95
Fluorine efficiency, %	98	99	98	99
Estimated y'	1.95	1.95	1.95	1.95
Estimated NTU (overall)	1.95	3.55	1.95	3.55

(1) These flows have been modified slightly to correspond to flows for DAP process.

(2) Based on 1/2 of 0.15 lb/ton permissible for entire process.

(3) Based on 1/4 of 0.15 lb/ton permissible for entire process.

TABLE 36

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR WET SCRUBBERS FOR GTSP PROCESS PLANTS

	MODERATE EFFICIENCY		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	40,900	65,300	Same	Same
°F	180	180		
SCFM	33,800	54,000		
Moisture Content, Vol. %	12.64	12.64		
Effluent Concentration Loading				
(ppm), Fluorine	195	195		
lb/hr, Particulate	116	185		
Cleaned Gas Flow				
ACFM	34,000	54,000	34,000	54,000
°F	100	100	100	100
SCFM	32,000	51,200	32,000	51,200
Moisture Content, Vol. %	6.47	6.47	6.47	6.47
Cleaned Gas Dust Loading				
(ppm), Fluorine	5.2	5.2	2.6	2.6
lb/hr, Particulate	5.4	8.8	5.4	8.8
Cleaning Efficiency, % Fluorine	98	98	99	99
(1) Gas Cleaning Device Cost	36,933	52,827	40,700	58,900
(2) Auxiliaries Cost	22,800	33,733	23,933	34,233
(a) Fan(s)	14,450*	25,300*	15,050*	25,200*
(b) Pump(s)	5,050*	7,650*	6,150*	8,150*
(c) Damper(s)	2,200*	2,750*	2,200*	2,750*
(d) Conditioning, Equipment	--	--	--	--
(e) Dust Disposal Equipment	--	--	--	--
(3) Installation Cost	86,197	116,647	90,457	122,657
(a) Engineering	6,200*	8,250*	6,200*	8,250*
(b) Foundations & Support	15,250*	22,100*	16,100*	23,300*
(c) Ductwork				
(d) Stack	16,650*	25,675*	16,650*	25,675*
(e) Electrical	4,850*	8,100*	5,350*	8,600*
(f) Piping	6,250*	9,050*	7,250*	10,550*
(g) Insulation				
(h) Painting	730*	1,185*	855*	1,375*
(i) Supervision				
(j) Startup	4,250*	5,750*	4,250*	5,750*
(k) Performance Test	2,150*	2,150*	2,150*	2,150*
(l) Other	13,750*	19,950*	14,800*	21,325*
(4) Total Cost	145,930	203,207	155,180	215,970

* Average of two bids. Third bidder did not itemize.

FIGURE 30

CAPITAL COST OF WET SCRUBBERS FOR GTSP PROCESS PLANTS

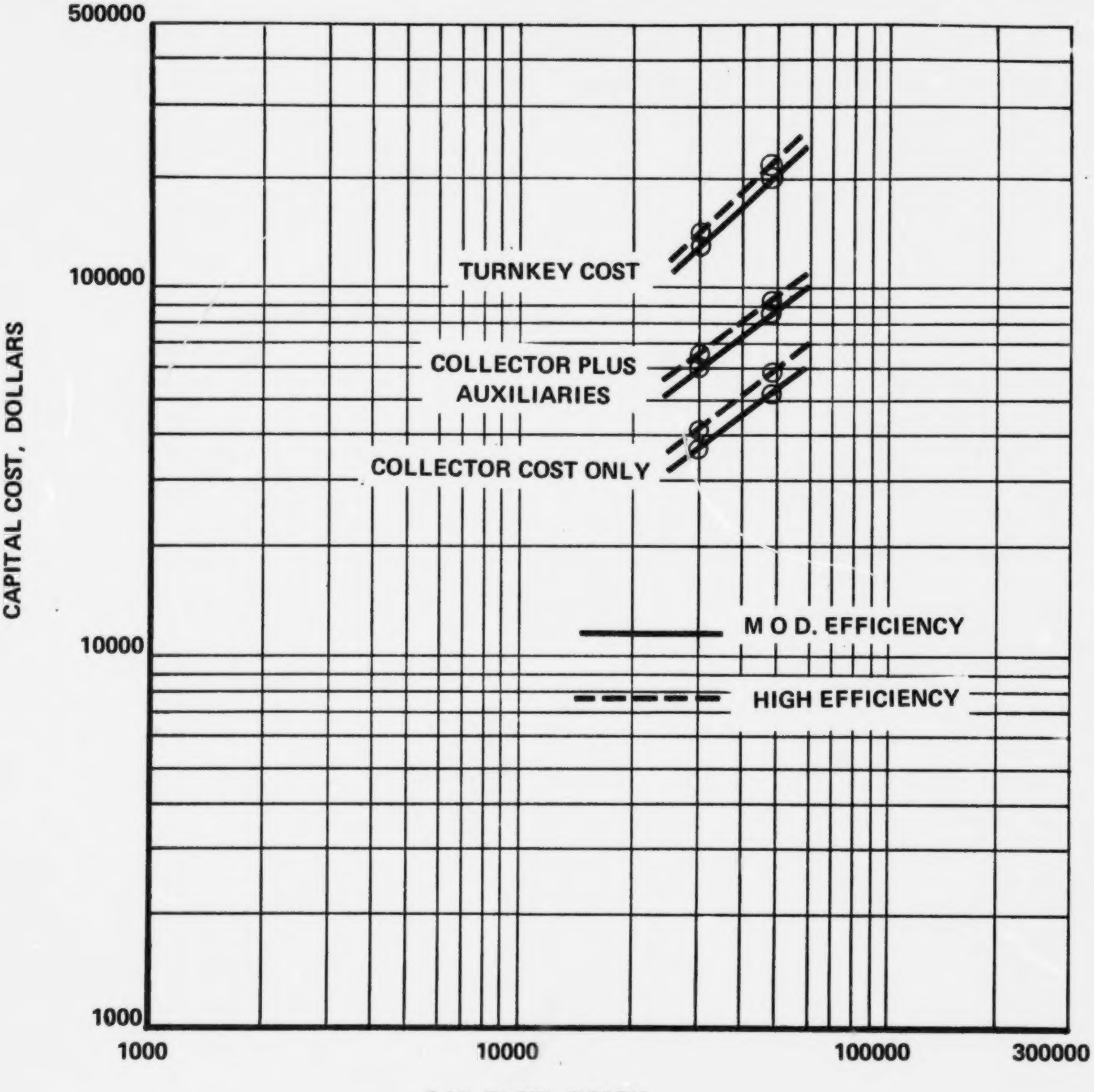


TABLE 37

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR GTSP PROCESS PLANTS (MOD. EFF. LEVEL)

Population – 5

Sample Size – 3

Capital Cost = $36,933

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$29,012	$44,855
75	21,585	52,282
90	10,156	63,710

Capital Cost = $52,827

Conf. Level, %	Lower Limit	Upper Limit
50	$48,883	$61,770
75	35,497	70,156
90	22,594	83,059

FIGURE 31

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR GTSP PROCESS PLANTS

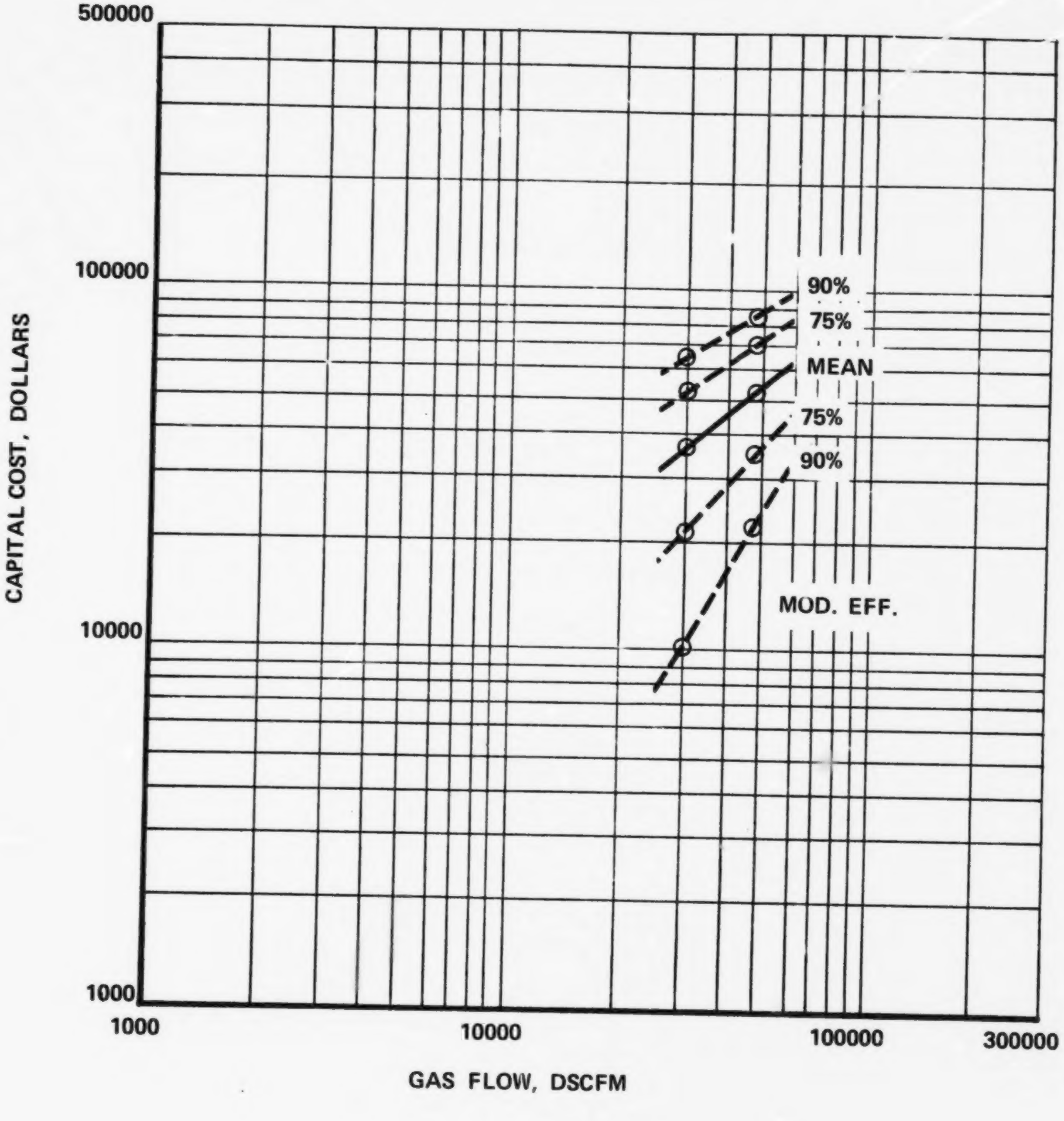


TABLE 38

CONFIDENCE LIMITS FOR INSTALLED COST OF WET SCRUBBERS FOR GTSP PROCESS PLANTS (MOD. EFF. LEVEL)

Population – 5 Sample Size – 3

Installed Cost = $145,930

Conf. Level, %	Installed Cost, Dollars Lower Limit	Upper Limit
50	$131,937	$159,923
75	118,818	173,042
90	98,630	193,230
95	79,962	211,898

Installed Cost = $203,207

Conf. Level, %	Lower Limit	Upper Limit
50	$191,793	$214,620
75	181,092	225,321
90	164,626	241,787
95	149,400	257,013

FIGURE 32

CONFIDENCE LIMITS FOR CAPITAL COST OF INSTALLED WET SCRUBBERS FOR GTSP PROCESS PLANTS

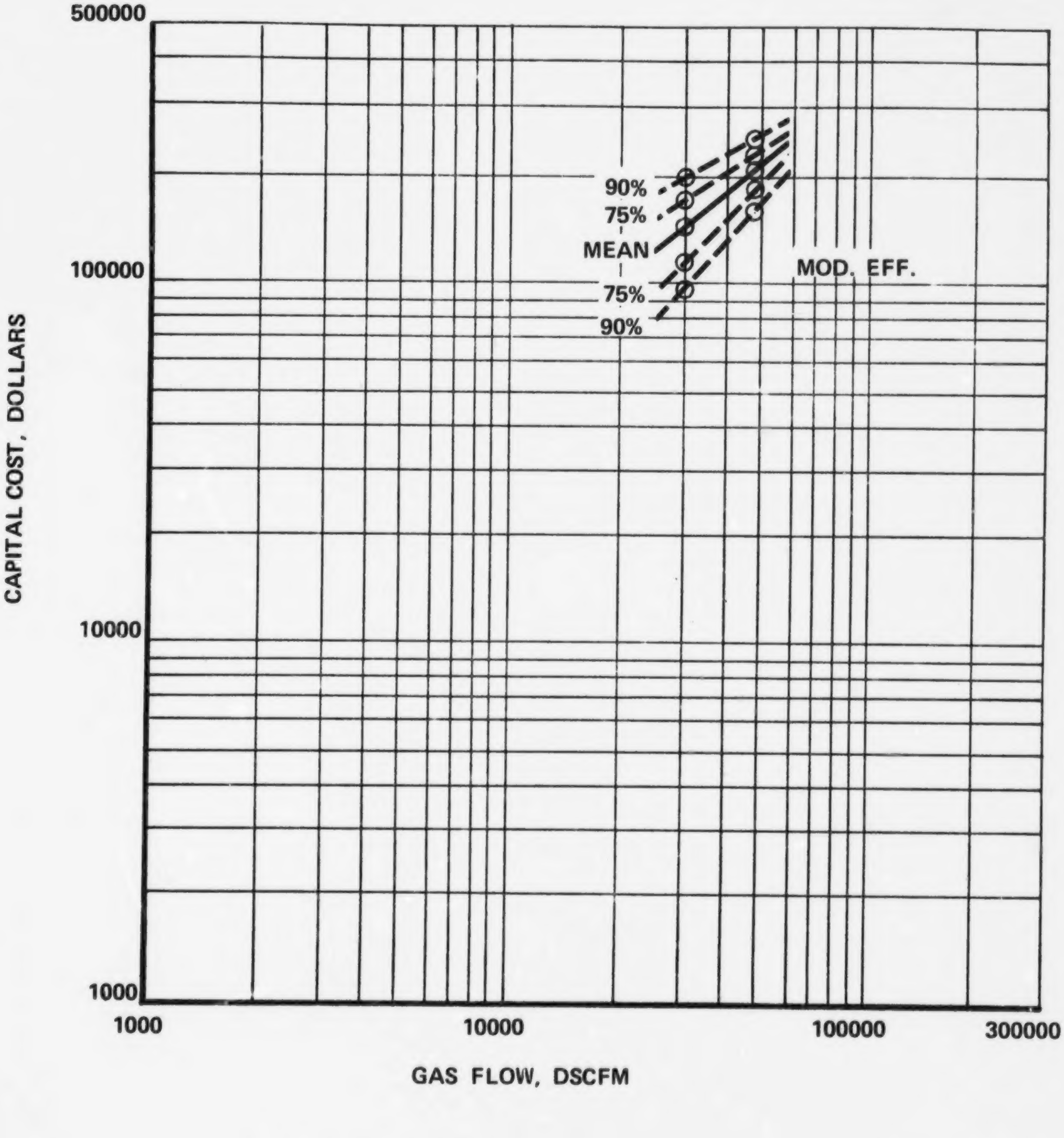


TABLE 39

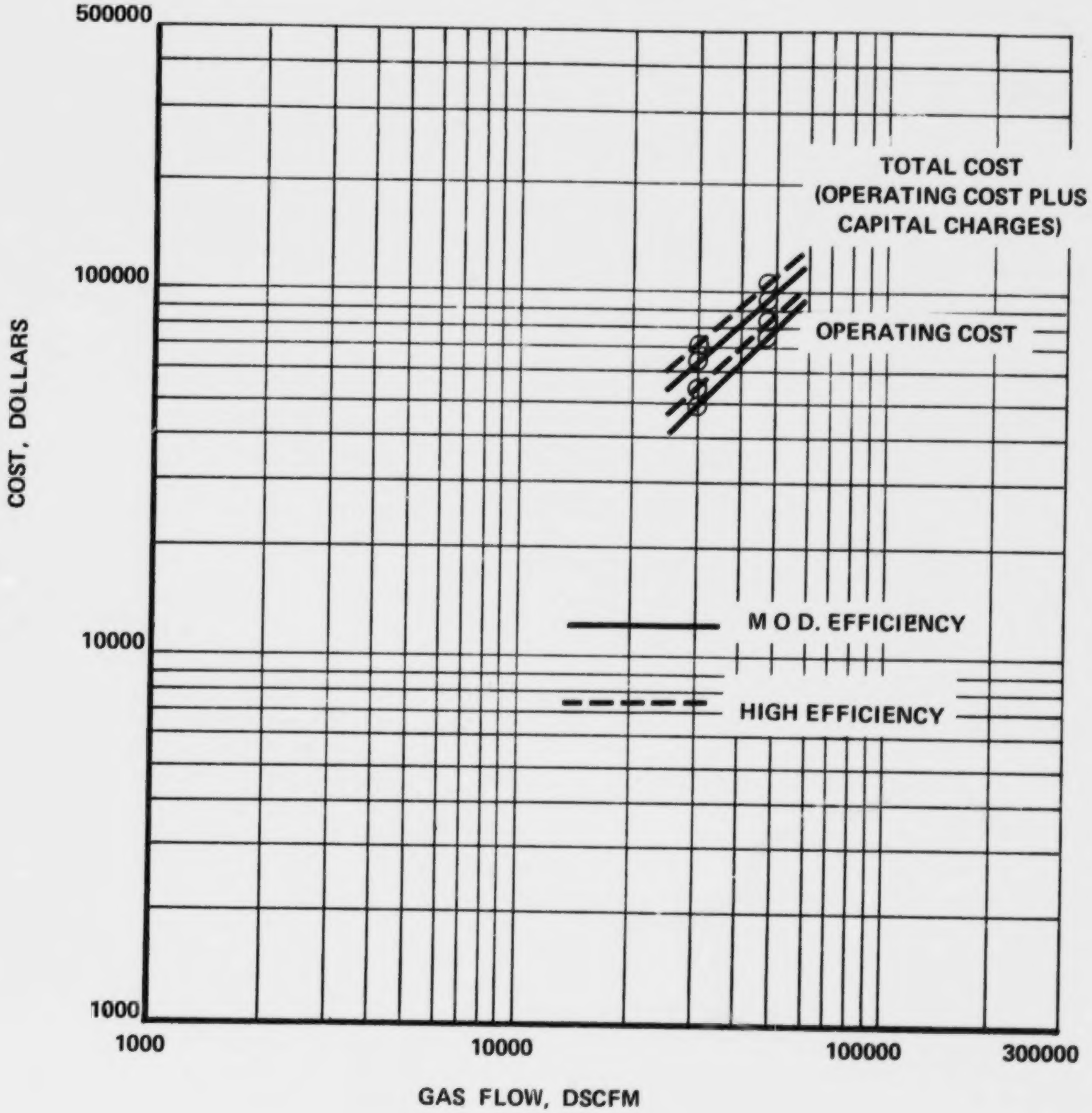
ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR WET SCRUBBERS FOR GTSP PROCESS PLANTS

Operating Cost Item	Unit Cost	MODERATE EFFICIENCY		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor		$ 1,950	$ 1,950	$ 1,950	$ 1,950
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		3,667	4,333	3,833	4,667
Replacement Parts					
Total Replacement Parts		4,017	5,800	4,950	7,433
Utilities					
Electric Power	$0.011/kw-hr	22,053	36,726	25,657	40,572
Fuel					
Recirculated Pond Water	$.05/M gal	14,776	23,660	16,136	25,420
Water (Make-up)	$.25/M gal	3,600*	5,400*	3,600*	5,400*
Chemicals, Specify					
Total Utilities		38,030	62,197	42,994	67,789
Total Direct Cost		47,664	74,280	53,727	81,842
Annualized Capital Charges		14,593	20,321	15,518	21,597
Total Annual Cost		62,257	94,601	69,245	103,439

*From one bidder only.

FIGURE 33

ANNUAL COST OF WET SCRUBBERS FOR GTSP PROCESS PLANTS



DIAMMONIUM PHOSPHATE (DAP)

Diammonium phosphate (DAP) is produced in plants similar to those used for the manufacture of granular triplesuperphosphate, and in fact the same plant is often used to produce these products alternatively.

DAP is made by first reacting anhydrous ammonia with phosphoric acid in a pre-reactor to form hot liquid DAP. This liquid is then pumped to an ammoniator-granulator drum, where it is further mixed with NH_3 and recycled solid material, and solidified. It is then dried, screened to size, cooled and sent to DAP product storage.

Acid produced by wet process phosphoric acid plants is used almost exclusively for the manufacture of diammonium phosphate fertilizers. Acid at 54% strength from the vacuum evaporation section of the plant is used for the initial preneutralization reactions, and acid at 30% strength is ordinarily blended to maintain the proper water level.(16)

DAP plants involve many of the same pollution considerations as WPPA, SPA, and GTSP processes treated previously in this report. However, in addition to the problems associated with absorbing gaseous fluoride compounds, the DAP process volatilizes a substantial amount of unreacted ammonia in the reactor and granulator-drier vessels. It is necessary to absorb the ammonia vapors in strongly acidic solutions for return to the reactor in order to minimize the loss of raw material and to avoid substantial air pollution by the ammonia discharge.

Process Description

Figure 34 is a process flow diagram(16) for a diammonium phosphate granulation plant. The following process description will make frequent reference to this diagram.

Anhydrous ammonia is obtained in bulk as a liquid under pressure. Liquid ammonia is ordinarily pumped from the storage vessel through a steam-jacketed coil in which the ammonia is vaporized. The ammonia vapor is bubbled through the liquid in a *preneutralization* reactor. Phosphoric acid at 54 wt.% is pumped through the reactor from the WPPA process. In addition, phosphoric acid at 26 to 30 wt.% which has been circulated through the reactor vent scrubber is also added to the preneutralization reactor.(7)

Substantial heat of reaction is evolved in the reactor, and some of the

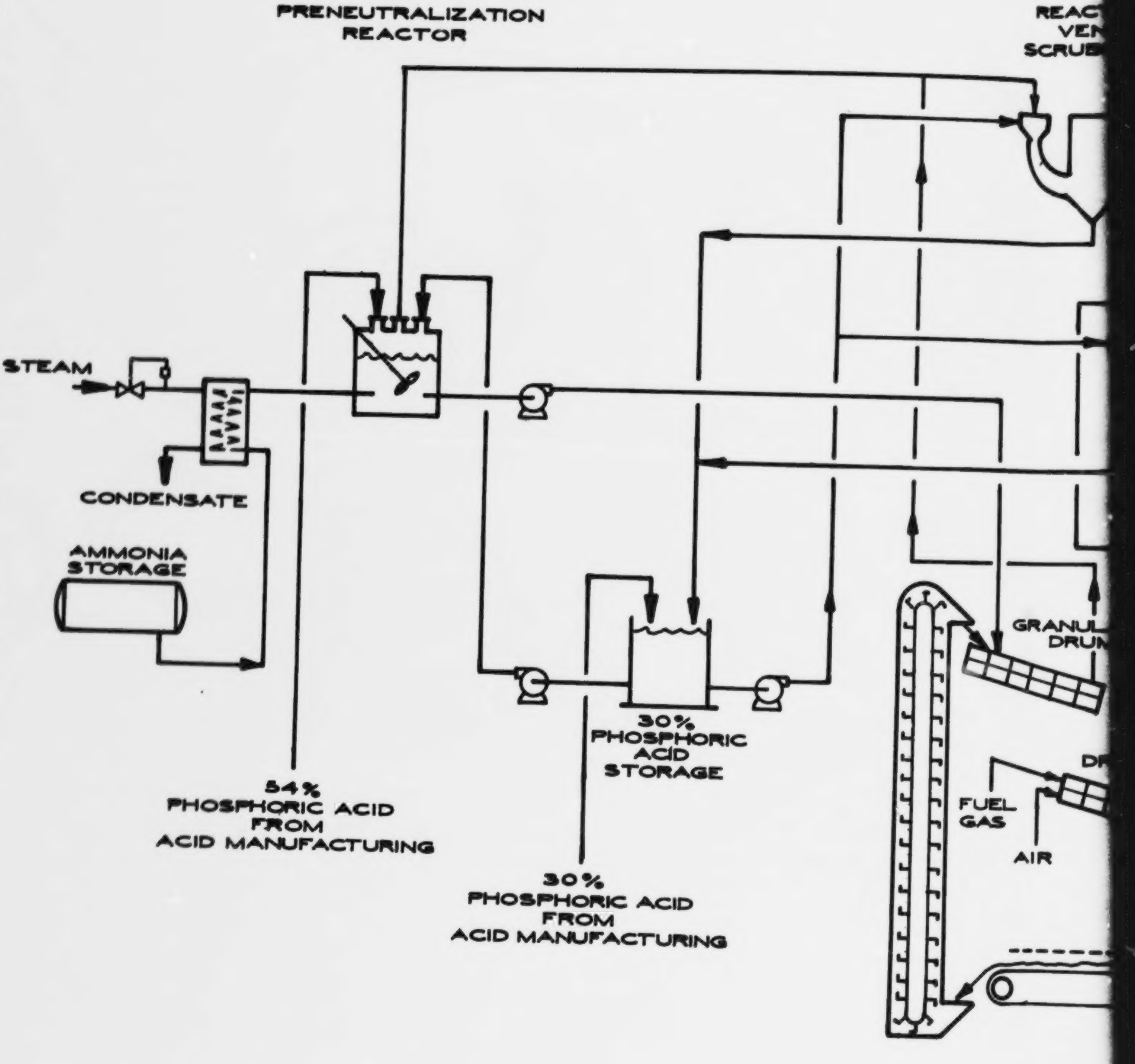
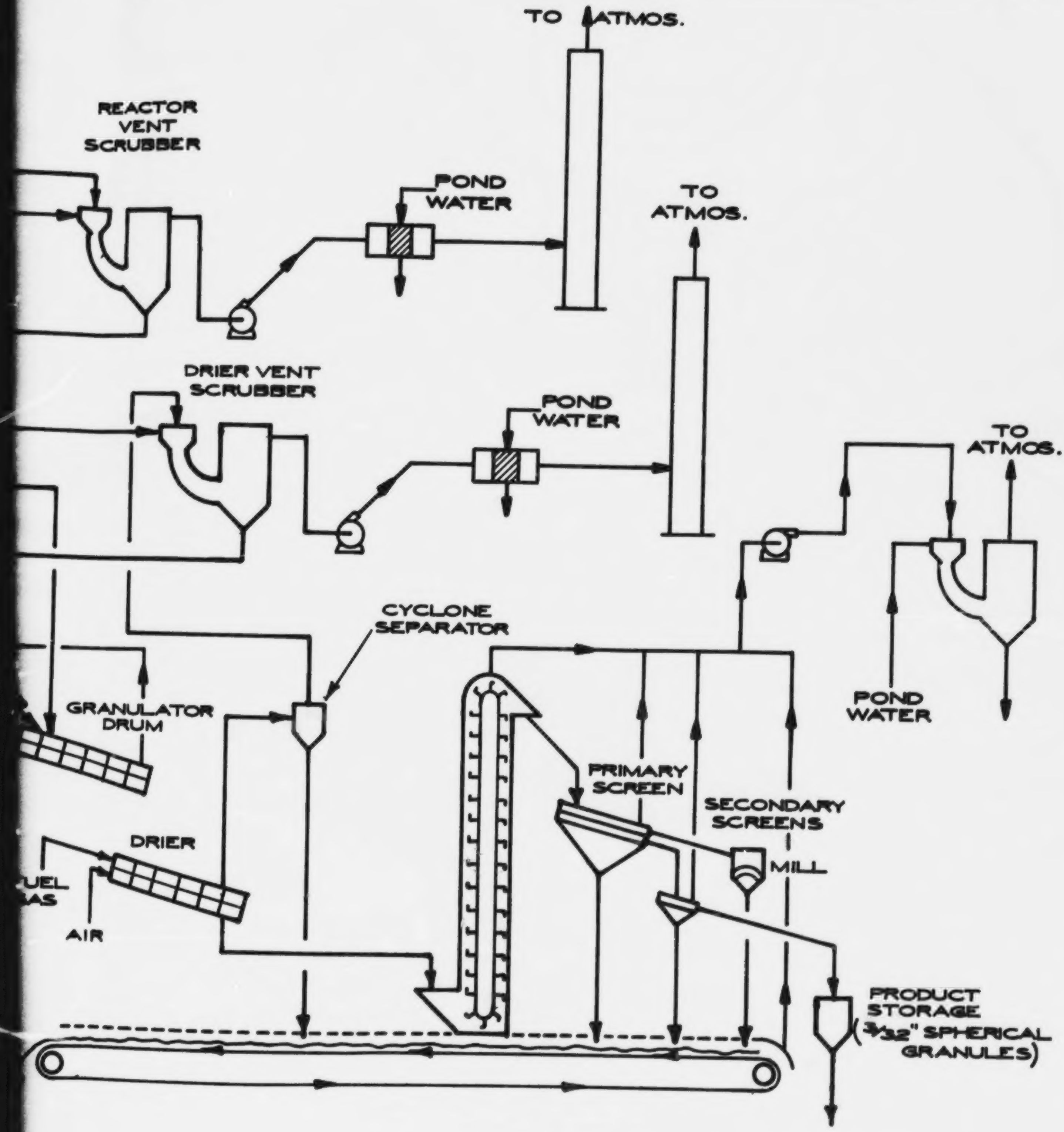

PRENEUTRALIZATION REACTOR
REAC
VEN
SCRUB
STEAM
CONDENSATE
AMMONIA STORAGE
54% PHOSPHORIC ACID FROM ACID MANUFACTURING
30% PHOSPHORIC ACID STORAGE
30% PHOSPHORIC ACID FROM ACID MANUFACTURING
GRANUL
DRUN
DR
FUEL GAS
AIR

FIGURE 34

FLOW DIAGRAM FOR
DIAMMONIUM PHOSPHATE (
GRANULATION PLANT


TO ATMOS.
REACTOR
VENT
SCRUBBER
POND
WATER
TO
ATMOS.
DRIER VENT
SCRUBBER
POND
WATER
TO
ATMOS.
CYCLONE
SEPARATOR
POND
WATER
GRANULATOR
DRUM
PRIMARY
SCREEN
SECONDARY
SCREENS
DRIER
MILL
UEL
AS
AIR
PRODUCT
STORAGE
(3/32" SPHERICAL
GRANULES)

URE 34

AGRAM FOR
PHOSPHATE (DAP)
TION PLANT

ammonia is discharged from the reactor as a vapor. Because there are no solid feed materials added to the reactor, it is possible to operate a nearly closed system from which only ammonia and steam, and little non-condensable gas such as air, must be discharged.(15)

The preneutralization reactor is operated with a somewhat lower mol ratio of ammonia to P_2O_5 than is required for the production of DAP, in order to prevent solidification of the entire mixture in this reactor. The resulting product is a slurry which can be pumped from the preneutralization reactor into the ammoniator-granulator. In the granulator, the slurry is exposed to additional ammonia vapor and solidification of the diammonium phosphate solid takes place.(6) In addition to the slurry which is fed to the granulator from the preneutralization reactor, a recycle of finely divided solid particles is added to the granulator from a recycle elevator as shown in Figure 25. The diammonium phosphate formed in the granulator and solidified from the slurry increases the particle size of the solids being recirculated through the granulator. The solid effluent from the granulator drum is dropped to a gas fired (or oil fired) rotary drier where excess moisture is driven off by direct contact with the flue gases. Flue gas from the drier is vented through a dry cyclone collector which collects the dust generated in the granulator-drier and returns it to a recycle belt. The granular solids discharged from the drier are elevated to a screening arrangement shown in the right hand side of Figure 25.

A primary screen removes undersized material from the flow stream and drops it onto the fines return conveyor. Oversized material is passed to a ball mill where it is ground to relatively small size and returned to the belt. The material passing the primary screen falls onto a secondary screen which rejects any fines which may have been carried over and passes the material of proper particle size through a rotary cooler into a product storage drum.(16)

All of the other solids, both oversized and undersized, are recycled by means of a single conveyor belt and elevator back to the granulator drum where they are trapped until the particle size has grown sufficiently large for them to be classified with finished product.

Particular attention must be given to the gas scrubber systems in the DAP process because they form an integral and necessary part of the process. Ordinarily, the vents from the preneutralization reactor and the granulator have a high ammonia content. In order to recover the ammonia contained in these gases, it is customary to circulate 20 to 30% acid (expressed as wt.% P_2O_5) through the scrubbers instead of pond water. The acid is circulated from a 30% acid storage tank as shown in Figure 34 through the primary scurbbing devices and the acid returned to the storage tank. Makeup acid from the reactor section of

Preceding page blank

the WPPA plant is added to the acid storage tank and 30% acid for addition to the preneutralization reactor is withdrawn from the tank.

CHEMISTRY OF THE PROCESS

In each of the processes described previously, it was necessary to oversimplify the reactions involving fluorapatite rock in order to illustrate the basic chemistry. In the case of the DAP process, the reactions are represented with good accuracy by the simple equation shown as follows:

$$\underset{\text{(Liquid)}}{H_3PO_4} + \underset{\text{(Gas)}}{2NH_3} \rightarrow \underset{\text{(Liquid)}}{(NH)_2\,HPO_4} - \underset{\Delta H_R \text{ at } 60^\circ F)}{81{,}500\ \text{BTU}} \qquad (26)$$

For purposes of this discussion, it may be assumed that the ammonia feed stock is substantially free of impurities. The phosphoric acid, on the other hand, contains some impurities which participate in side reactions.

The acid composition given in Table 7 may be used for the development of hypothetical process emissions, although it should be borne in mind that the use of 30% acid or feed grade WPPA acid in addition to the 54% acid results in bringing more fluorine into the reaction than indicated by the Table 7 composition.

The side reactions involve the formation of ammonium sulfate, ammonium fluoride, and ammonium fluorosilicate. Also, some of the fluorine is liberated as silicon tetrafluoride and discharged from the reactor with the ammonia gas. It should be noted that the *product weight*, rather than the P_2O_5 content of the product, is the usual basis for rating the DAP plant. It is apparent from Table 40 that a substantial amount of the water present in the phosphoric acid is removed in the drying process.

NATURE OF THE GASEOUS DISCHARGE

The DAP process has two major sources of gaseous discharge and two dust sources which may be significant from a pollution standpoint:

Gaseous

1. Reactor and granulator ventilating gas

2. Drier flue gas

Dust

1. Cooler air discharge*
2. Transfer points and screen ventilation

In this section, all three of these sources will be considered, and pollution control equipment applicable to each source will be specified, and cost data will be obtained for each.

Reactor

The reactor or preneutralizer is a vessel into which about 70% of the ammonia and all of the phosphoric acid are introduced. In some plants, 93 wt.% H_2SO_4 is introduced to control final analysis of the product.

The reaction generates a substantial amount of heat which raises the temperature of the reactants and results in the discharge of large quantities of water vapor, unreacted ammonia and SiF_4. In order to assure that there will be no leakage of ammonia and phosphoric acid from the mixing vessel, the vessel is well ventilated with outside air. The ventilating air flow rate is a critical factor in the design of subsequent gas treating equipment, and this is set on the basis of mechanical factors in the reactor design, rather than by any fundamental requirement of the process. In theory, the reactor could be designed to avoid any ventilating air flow and the system run without discharge to the atmosphere.

In practice, ventilating air flow rates of around 2,500 SCFM through the reactor are common.(15) The ventilating air serves to remove some of the heat of reaction, and also increases the flow of ammonia out of the reactor vessel. The ammonia "lost" from the reactor at this point is captured in the preneutralizer scrubber through which 20 to 30% concentration acid is circulated. Concentrations around 26% are preferred to keep the product in the MAP range.

The rate of gas flow through the preneutralizer is established by the inward air leakage required to prevent loss of ammonia and fluorine-containing gases from the preneutralizer. Inward air leakage rates on the order of 2,000 SCFM are reasonable for reactors of commercial design. To this air stream is added ammonia, water vapor, and minor components of the DAP slurry in

---

*In some cases NH_3 losses have been observed where inadequate operation of the dryer shifts losses to the cooler.

TABLE 40

APPROXIMATE MATERIAL BALANCE OF 1,000 TON/DAY DAP PROCESS

	Acid Feed (@ 40% P_2O_5)		Ammonia Feed		DAP Product @ 1 wt.% Moisture	
	Ton/Day	Lb/hr	Ton/Day	Lb/Hr	Ton/Day	Lb/Hr
NH_3	–	–	240	20,000	–	–
P_2O_5	500	41,690	–	–	–	–
$(NH_4)_2HPO_4$	–	–	–	–	930	78,500
F	14.3	1,190	–	–	12.2	1,015
SO_3	15.8	1,315	–	–	15.8	1,315
Al_2O_3	18.6	1,545	–	–	18.6	1,545
Fe_2O_3	7.4	620	–	–	7.4	620
Other	3.7	310	–	–	3.7	310
Water	690.2	57,500	–	–	9.8(1)	810
Total	1,250.0	104,170	240	20,000	997.5	84.115

(1) Water removed is (690.2 Ton/Day total water – 190 Ton/Day combined water in phosphoric acid – 9.8 Ton/Day in product = 490.2 Ton/Day).

portion to the vapor pressure at the reaction temperature. Water vapor will move toward equilibrium between the gas stream and a liquid phase which is less than 60% water. The gas stream will never contain more than about 60% saturated with respect to liquid water in the reactor, but will be close to saturation at the exhaust temperature of 170 to 180° F.

At a discharge temperature of 180° F, the estimated composition of the effluent gases for a reactor with 2,000 SCFM in-leakage of air is given in Table 41. The fluorine content of this gas stream, either as particulate or vaporized components has in the past been of little concern as it leaves the reactor. The ammonia content is too high for discharge into the atmosphere for reasons of economy and pollution control, and it is necessary to contact the ammonia-bearing gas stream with relatively concentrated phosphoric acid in order to minimize ammonia loss. The contracting process serves to recover the ammonia effectively, but in doing so results in volatilization of fluorine compounds from the absorbing liquor.(6) Today, the fluorine stripping effect is sufficient to warrant the addition of a tail gas scrubber after the primary absorber. In the hypothetical process selected for study here, it will be assumed that 30% acid is used for the scrubbing liquid and that the fluorine concentration in the tail gas is significant.

GRANULATOR

A second significant source of emissions from the process is the granulator.(8) Here, a solids recycle material consisting of under-sized solid product is fed to the granulator from a bucket elevator or other loading device. In addition, the slurry from the preneutralizer and the remaining ammonia are added and final solidification of the DAP takes place. As in the case of the preneutralization reactor, the granulator might be operated without any net flow of gas, except that it is mechanically very difficult to maintain a sealed rotary unit into which solids are fed. For this reason, granulators are designed with a substantial in-leakage of air around the inlet and outlet connections. The ventilating air flow purges a substantial amount of ammonia from the granulator along with some water vapor and DAP dust.

There is very little evolution of fluorine-containing gases from the granulator because of the high concentration of ammonia in the vapor phase. Again, the recovery of ammonia from the ventilating gas is the primary concern in the treatment of the off gases from the granulator. Substantial concentrations of particulate matter in the DAP pose a potential fluoride emission problem and particulate collection with return of the DAP

TABLE 41

COMBINED VENTS FROM REACTOR AND GRANULATOR

	Reactor	Granulator	Total to Scrubber	Total from Scrubber
Temp, ° F	190	150	170	135
Moisture, vol %	50	39	40	40.8
Moisture, lb/lbs Gas	0.625	0.39	0.415	0.430
Gas Composition, SCFM				
Water	2,000	11,250	13,250	13,750
Air	2,000	18,000	20,000	20,000
Ammonia	20	200	220(2)	12(1)
	4,020	29,450	33,470	33,762

(1) Based on 3 lb/ton NH_3 feed.
(2) Based on 11.2 ton/hr in.

particulates to the preneutralization reactor along with the phosphoric acid scrubbing liquid is essential to the process. An estimate of the composition of gas from the granulator is given in Table 41. The rotary drier removes the free water from the DAP by direct contact with flue gases. The flue gas generation rate may be calculated approximately by heat balance of the drier. The heat input must be sufficient to vaporize the difference between the water content of the dry product and the water input with both phosphoric acid streams. Little heat is added to the solid product because it enters the dryer combination at nearly the exit temperature.

A rough heat balance of the dryer for a 500 Ton/day plant is shown in Table 42, and the calculated gas flow on the heat balance is used in specifying the abatement equipment at the end of this section.

At the dryer outlet, there must be some provision for in-leakage of ventilating air to prevent loss of flue gas and product dust. The flue gas-steam combination produced in the dryer is ordinarily vented through a mechanical collector and serves to collect particulate solids for return to the process. The exhaust gas from the cyclone is scrubbed either in combination with the gas ventilated through the preneutralization reactor, or in a separate scrubber. The two services may be combined if a common phosphoric acid circulating stream is used for both of them. In the flow scheme shown in Figure 34, the streams are treated separately, and in the hypothetical plant for which specifications and prices are given, this is the case.

COOLER

The final emission sources which must be treated in a DAP plant are the air used for cooling the solid product and ventilating the solids transfer points. The product cooler reduces the temperature from 190° F or thereabouts to about 110° F by direct contact of cooling air with the solids in a rotary drum. The air flow requirement can again be calculated on an approximate basis by a heat balance of the process. There is little in the way of volatile material given off in the cooler as the temperature is reduced and the vapor pressure of each of the species decreases. However, there is a substantial dusting problem due to the mechanical handling of the granular solids. The principal pollution control problem associated with the cooler is the recovery of this particulate matter. Because the ammonia has been stripped out of the solid at this point, it is not necessary to use acidic water for scrubbing, and pond water may be used in this scrubber for convenience.

TABLE 42

ROUGH HEAT BALANCE FOR DAP DRYER FOR 1,000 TON/DAY PROCESS

	Flow SCFM	Flow lb/hr	Temp., °F	Heat Content BTU/lb	Heat Content BTU/hr
Into Dryer					
Wet DAP	–	124,905	200	–	8.12
DAP	–	(83,305)	200	(32.5)	(2.71)
Water	–	(41,600)	200	(130)	(5.41)
Recycle Solids	–	420,725	200	33.5	14.10
Air to Burner(1)	9,000	41,000	70	0	0
Fuel to Burner(2)	900	2,280	70	0	0
Tempering Air(3)	21,900	100,000	70	0	0
Dilution Air(4)	–	37,000	–	0	0
Total Gases	30,900	143,280			22.22
Out of Dryer					
DAP Product	–	84,145	200	33.5	2.82
DAP	–	(83,305)	200	(32.5)	(2.71)
Water	–	(840)	200	(130)	(0.11)
Recycle Solids	–	420,725	200	–	14.10
Flue Gas	46,000	184,040	180	293	53.80
N_2	(24,480)	(108,500)	180		
O_2	(4,620)	(23,400)	180	27.5	(3.80)
CO_2	(900)	(6,240)	180		
H_2O	(16,100)	(45,900)	180	(1,090)	(50.0)
					70.70

Burner Duty = 70.72 – 22.22 = 48.50 X 10^6 BTU/hr.

(1) Based on 10/1 air to fuel ratio for gas firing
(2) Based on 900 BTU/SCF lower heating value
(3) Tempering air to produce 1,500° F temperature to dryer
(4) Dilution air to reduce outlet temperature to 180° F

TRANSFER POINTS AND SCREEN VENTILATION

The DAP granulation process, as illustrated in Figure 34 involves the handling and screening of a great deal of granular solid product and recycle solids. At each transfer point (i.e., elevator, belt conveyor, screen or mill discharge) there is some generation and dispersion of product dust. This must be retained within the process by exhausting air from exclosures around the transfer points. This dust contains a small amount of fluorine as an impurity, but consists mainly of DAP product.

Recovery of the dust and return to the process offers some potential for improved product yield, and phosphoric acid may be used as the scrubbing medium. The dust may, however, be collected in scrubbers circulating pond water, which discharges into the gypsum pond.

APPLICABLE ABATEMENT EQUIPMENT

This discussion deals with the wet scrubbing equipment conventionally applied for removal of ammonia, fluorides and particulate dusts from effluent gas streams generated in DAP plants. These scrubbers are designed for a variety of functions which include:

Ammonia Recovery
Particulate Collection (and sometimes recovery)
Fluorine Removal

These functions require a complex arrangement of the scrubbing equipment.

The three major process sources in DAP plants are treated exclusively by wet scrubbers. Several circumstances contribute to the selection of wet scrubbers over other types of pollution control devices for these services. The necessity for removal of gaseous ammonia and fluoride compounds from the gas stream is a significant factor in DAP process plants, although the emphasis here is more on control of particulate emissions than is the case for any of the phosphate processes described in earlier sections. A second contributing factor is the relatively high humidity of the gas streams, all of which originate in wet reactors, driers, or other pieces of process equipment from which water vapor emission is significant. The high concentration of water in the gas streams poses problems in the use of fabric collectors, and, to a lesser extent, in the use of mechanical or electrostatic collectors.

Finally, the ready availability of pond water as a scrubbing medium, and the gypsum pond as a settling basin for collected solids is a significant advantage for wet scrubbers.

The combination of requirements for particulate collection and gas absorption for NH_3 recovery and fluorine emission control permits the application of a wide variety of scrubber types for DAP service. The devices most used are:

1. Two-stage wet cyclonic
2. Venturi-cyclone* scrubbers
3. Cross-flow packed scrubbers

Figure 35 shows a schematic drawing of each of these three types. These scrubbers are ordinarily used as follows:

	Primary Scrubber	Secondary Scrubber
Reactor granulator	Venturi-Cyclone or Two-Stage Cyclonic	Cyclonic or Packed Cross-flow
Dryer		
Cooler + Transfer points	Venturi Cyclone or Two-Stage Cyclonic	

The two-stage wet cyclonic scrubbers are relatively simple in design and construction, and have relatively low pressure drop characteristics. They are well suited for the collection of particulate greater than about 3μ in size, and for gas absorption. They are less efficient as absorption devices than are the scrubbers utilizing wetted packings.

---

*"Venturi-cyclone", rather than the more common "Venturi-cyclonic" is used here to distinguish a scrubber in which the "cyclone" is used for entrainment separation only. "Cyclonic" is reserved for scrubbers using the cyclonic scrubbing principal.

The Venturi cyclone, operating in the range of 15 to 30 inches water column pressure drop, is perhaps the best particulate collection scrubber of the three and also serves adequately for ammonia absorption, especially when a packed section is installed after the primary Venturi cyclone collector, or when water sprays are used in the entrainment separator to enhance the gas absorption characteristics of the unit.

Typically, a 30 in. w.c. pressure drop Venturi can be used to collect not only the dusts, but also the condensation products such as ammonium fluoride in the sub-micron size range, formed by gas phase reactors.

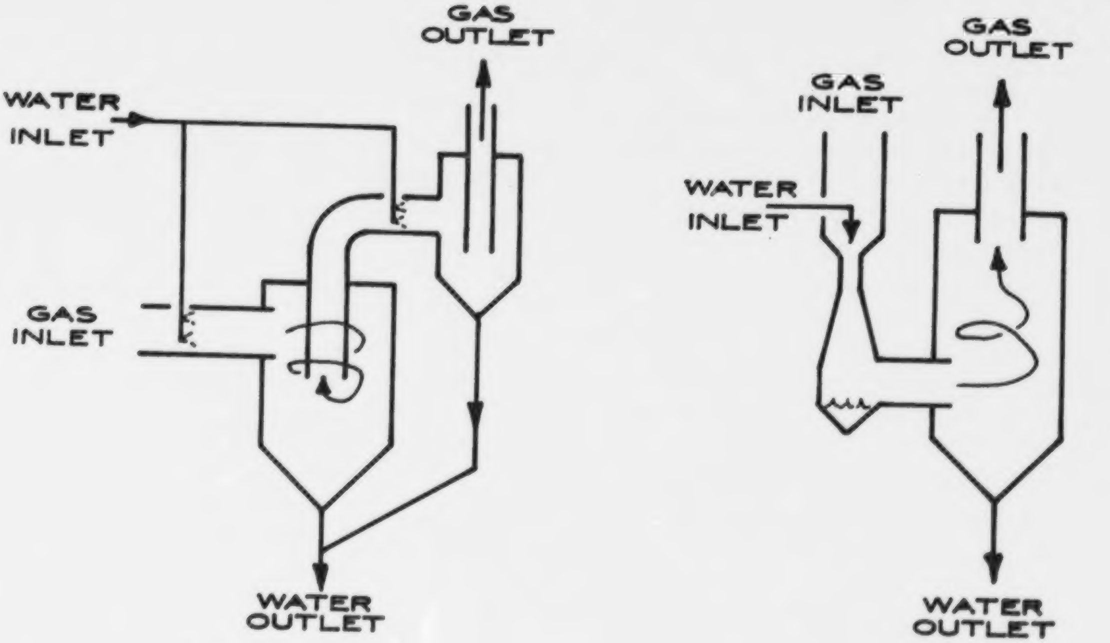
The cross-flow packed scrubber is the best of the three devices from a gas absorption standpoint, but is less effective for the collection of particulate matter, and for this reason is used as a "tail gas" scrubber following a Venturi or two-stage cyclonic. The packed scrubbers are seldom used as primary scrubbers because of the tendency to plug with gelatinous silicon or DAP deposits.

In order to generate a reasonably complete picture of the costs of these scrubbers in the various services, a generalized pattern has been followed in writing specifications. The equipment is presumed to be installed on a 1,000 T/D DAP process (500 T/D P_2O_5 content) and individual scrubbers specified as follows:

Reactor-Granulator – 20,000 DSCFM
    Venturi-Cyclone
    Two-Stage Cyclonic
Dryer – 30,000 DSCFM
    Venturi-Cyclone
    Two-Stage Cyclonic
Tail Gas Service

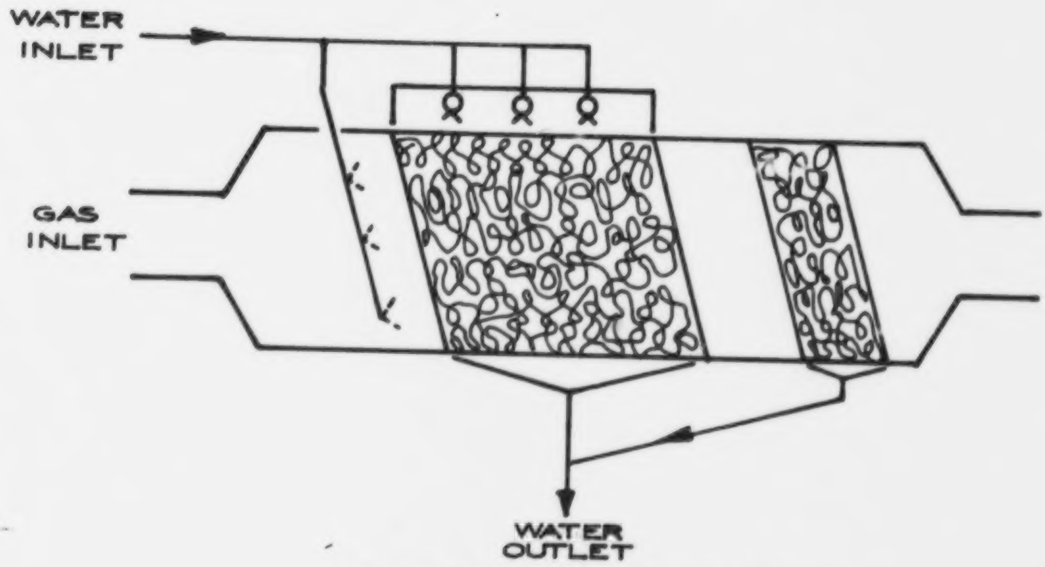
The particulate collection scrubber should always be the "upstream" member of a two scrubber combination. That is, a two-stage wet cyclonic or Venturi-cyclone would always precede a cross-flow packed scrubber. It is difficult to envision circumstances that would lead to the installation of a cross-flow packed scrubber ahead of a two-stage wet cyclone. Figure 36 illustrates those combinations which are reasonable from the standpoint of performing the solid collection in the scrubber most suited for handling solids and the gas absorption in subsequent scrubbing steps.

As in each of the previous cases, the equilibrium between the fluorine-containing gas, SiF_4 and HF, and H_2SiF_6 in the pond water is o


GAS OUTLET
WATER INLET
GAS INLET
WATER OUTLET
GAS OUTLET
GAS INLET
WATER INLET
WATER OUTLET

TWO-STAGE CYCLONIC SCRUBBER

VENTURI CYCLONIC SCRUBBER


WATER INLET
GAS INLET
WATER OUTLET

CROSS-FLOW PACKED SCRUBBER

FIGURE 35

SCHEMATIC DRAWING OF SCRUBBERS USED IN DAP TAIL GAS TREATING considerable importance. The equilibrium vapor pressures of these gases given in the discussion of the WPPA process indicate that the gas stream must be relatively cool in order for low enough values of fluorine in the ventilating gas to meet the present Florida law to be realized. Ordinarily, gases entering a scrubber are at a higher temperature and lower moisture content than those exiting, due to the saturation and cooling of the gas which takes place in the scrubber. In the DAP plant, the moisture content of gases vented to the tail gas scrubber is too high for evaporative cooling to take place, and some condensation of moisture from the gas must take place in the scrubber. This requires that a substantially higher flow rate of scrubber liquor per unit of gas must be passed through the scrubber than would be the case for cooling of the gas by evaporation of moisture from the scrubbing liquor.

One other problem is peculiar to DAP scrubbing among the other phosphate rock processing pollution abatement processes. Because recovery of ammonia is of critical importance in the process economics, the scrubbers may be operated at temperatures as high as 185°F for contacting the ammonia-containing vent gases with 30% phosphoric acid. Many of the materials of construction most desirable from a corrosion standpoint (PVC, FRP and rubber) have substantially decreased physical properties in this temperature range, and pose some specific problems in fabrication of large scrubbers of these materials. In particular, PVC is unlikely to be acceptable in this service because of the high temperatures.

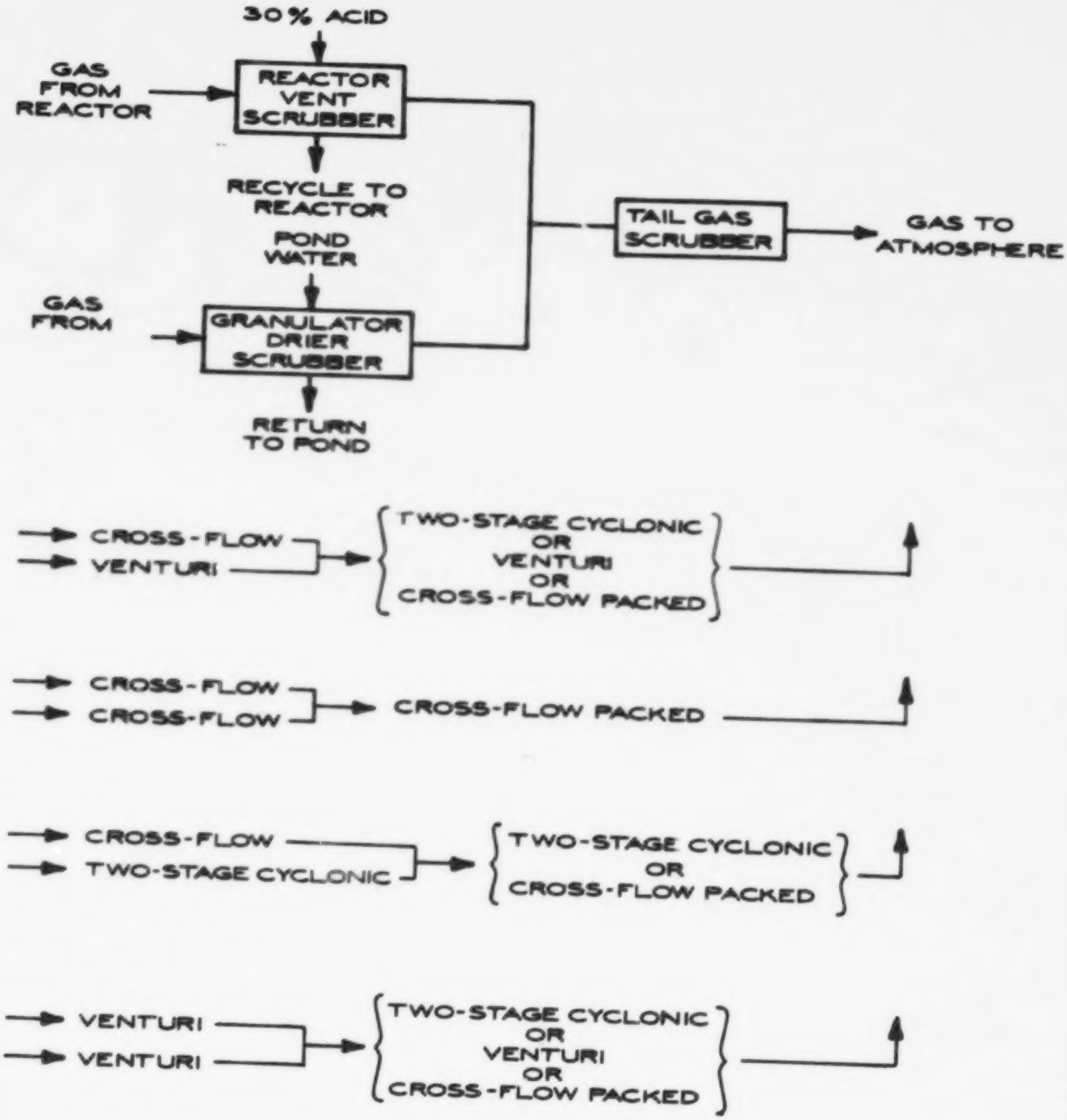

30% ACID
GAS FROM REACTOR
REACTOR VENT SCRUBBER
RECYCLE TO REACTOR
POND WATER
GAS FROM
GRANULATOR DRIER SCRUBBER
RETURN TO POND
TAIL GAS SCRUBBER
GAS TO ATMOSPHERE
CROSS-FLOW
VENTURI
TWO-STAGE CYCLONIC
OR
VENTURI
OR
CROSS-FLOW PACKED
CROSS-FLOW
CROSS-FLOW
CROSS-FLOW PACKED
CROSS-FLOW
TWO-STAGE CYCLONIC
TWO-STAGE CYCLONIC
OR
CROSS-FLOW PACKED
VENTURI
VENTURI
TWO-STAGE CYCLONIC
OR
VENTURI
OR
CROSS-FLOW PACKED

FIGURE 36

ACCEPTABLE SCRUBBER COMBINATIONS FOR DAP PROCESS PLANTS

TABLE 43

PROCESS DESCRIPTION FOR WET SCRUBBERS FOR DAP PROCESS

This specification describes three types of scrubbers and three process gas streams. There is some degree of interchangeability of the scrubbers specified.

The DRYER is to be treated for particulate and ammonia removal by either a Venturi-cyclone or a two-stage wet cyclonic scrubber. Thirty percent phosphoric acid is used as the scrubbing medium in these scrubbers. Either is followed by a crossflow packed scrubber using water as the scrubbing medium for fluorine removal. These scrubbers should be identical to those specified for the Reactor-Granulator vent, except for size. The cooler and dryer are to be considered interchangeable applications.

Two efficiency levels are specified for each scrubber. A separate set of process conditions is specified for each size.

The pond water to be used for scrubbing has characteristics as follows:

	Design	*Min.*	*Max.*
Pond Water, pH	*2.0*	*1.2*	*2.2*
Temperature, ^oF	*80**	*55*	*88*
SO_4, Wt.%	*0.15*		
P_2O_5, Wt.%	*0.7*		
H_2SiF_6, Wt.%	*0.63*	*0.25*	*1.0*
Fluorine, Wt.%	*0.5*	*0.2*	*0.8*

**NOTE: Pond water temperature may be considerably higher than the values given here. Design temperatures of 90^oF+ may be specified, and correspondingly high gaseous fluorine limits may be expected.*

The scrubber is required to produce the specified performance when operating with water at the "design" conditions. The vendor shall specify the water flow rate to be used, and that this flow rate is adequate for his scrubber design.

Materials of construction shall be limited to the following, which shall be selected for hydrofluoric acid service:

Rubber-Lined Steel

FRP (Dynel Lined)

No metal parts shall be used where exposure to process gas or pond water may be significant.

MECHANICAL

A. Two-Stage Wet Cyclonic

The scrubber shall be equipped with a cone bottom, bottom tangential gas inlet, two stages of sprays which can be removed for cleaning of spray nozzles while the unit is operating, and a top gas outlet.

A fan capable of overcoming the pressure drop across the scrubber stages and ductwork shall be supplied as an auxiliary and shall be located adjacent to the scrubbers. The fan shall operate on the discharge side of the cyclone.

A separate hotwell with a barometric leg seal must be supplied, and should be enclosed to minimize fluorine losses.

The scrubber system shall be furnished complete with all interconnecting ductwork from the scrubber to the fan and from the fan to the stack. A return pump, 50 ft. stack and barometric damper to minimize pressure fluctuations at the dryer shall be provided.

The scrubber and auxiliaries are to be located outside, adjacent to the dryer.

B. Venturi-Cyclone

The Venturi cyclone scrubber shall have a pressure drop of 15-30 inch water column and shall be designed to have no spray nozzles in the Venturi section. All interconnecting ductwork, fan, return pump, etc. similar to that outlined under the Two-Stage Wet Cyclone shall be provided. A variable throat may be substituted for the barometric damper to maintain the dryer peressure constant.

C. Cross-Flow Packed

The cross-flow tail gas scrubber, if required, should be placed in series after either the two stage cyclone or the Venturi cyclone. The fan shall be placed between the primary scrubbers and the cross-flow tail gas scrubber.

The cross-flow scrubber should be designed to treat the gas stream in a single packed bed of sufficient depth to obtain the required removal efficiency. Top irrigation as well as a front face spray of the wetted bed should be used. High top liquid irrigation rates should be used in the front packed section to prevent solids deposition.

A dry packed bed shall serve as an entrainment separator. Top water sprays shall be furnished in the entrainment separation section for intermittent wash-down of the packing. The scrubber sump bottom should be sloped to prevent solids build-up and should contain baffles to prevent gas by-pass of the packed bed.

When the cross-flow packed scrubber is required all additional interconnecting ductwork, return pump, etc., shall be provided in addition to that outlined in the Two-Stage Wet Cyclone and Venturi Cyclone sections to comprise a complete system.

TABLE 44

*OPERATING CONDITIONS FOR PRIMARY SCRUBBERS FOR DAP PROCESS DRYER VENTS**

Plant Capacity, Ton/Day DAP	*1,000*	*1,600*
Plant Capacity, Ton/Day P_2O_5	*500*	*800*
Process Weight, Ton/Hr	*42*	*66.7*
Gas to Scrubber		
Flow, DSCFM	*30,000*	*48,000*
Flow, SCFM	*46,100*	*74,000*
Flow, ACFM	*55,700*	*89,000*
Temp., ^oF	*180*	*180*
Moisture, vol.%	*35*	*35*
Particulate, gr/SCF	*0.5*	*0.5*
Particulate, Lb/Hr	*200*	*320*
Ammonia, Lb/Ton P_2O_5	*8.0*	*8.0*
Ammonia, Lb/Hr	*167*	*267*
Ammonia, ppm	*1,350*	*1,350*
Gas from Scrubber		
Flow, DSCFM	*30,000*	*48,000*
Flow, SCFM	*46,000*	*75,000*
Flow, ACFM	*53,500*	*85,500*
Temp., ^oF	*164*	*164*
Moisture, vol.%	*36*	*36*
Particulate, gr/SCF	*0.01*	*0.01*
Particulate, Lb/Hr	*4.0*	*6.4*
Ammonia, Lb/Ton P_2O_5	*0.08*	*0.08*
Ammonia, Lb/Hr	*1.67*	*2.67*
Ammonia, ppm	*13.4*	*13.4*
Particulate Efficiency, wt.%	*98*	*98*
Ammonia Efficiency, %	*99*	*99*

**These are considered interchangeable with cooler vent scrubbers.*

TABLE 45

OPERATING CONDITIONS FOR SECONDARY SCRUBBERS FOR DAP PROCESS DRYER VENTS*

Plant Capacity, Ton/Day DAP	1,000	1,600
Plant Capacity, Ton/Day P_2O_5	500	800
Process Weight, Ton/Hr	42	66.7
Gas to Scrubber		
Flow, DSCFM	30,000	48,000
Flow, SCFM	46,800	75,000
Flow, ACFM	53,500	85,500
Temp., °F	164	164
Moisture, vol.%	36	36
Fluorine, ppm as F-	15	15
Fluorine, Lb/Hr	2.1	3.35
Gas from Scrubber		
Flow, DSCFM	30,000	48,000
Flow, SCFM	31,900	51,000
Flow, ACFM	33,700	54,000
Temp., °F	100	100
Moisture, vol.%	7	7
	Medium Efficiency Case	
Fluorine, ppm as F-	4.1	4.1
Fluorine, Lb/Hr(1)	0.415	0.67
Fluorine removal, %	80	80
Estimated y', ppm	1.95	1.95
Estimated NTU	1.80	1.80
	High Efficiency Case	
Fluoride, ppm as F-	3.25	3.25
Fluoride, Lb/Hr(2)	0.31	0
Fluorine removal, %	85	85
Estimated y'	1.95	1.95
Estimated NTU	2.8	2.8

*These are considered interchangeable with cooler vent scrubbers.

(1) Based on 1/3 of 0.06 Lb F/Ton P_2O_5 allowable under present Florida law.

(2) Based on 1/4 of 0.06 Lb F/Ton.

SPECIFICATIONS AND COSTS

Tables 41 - 43 contain specifications for scrubbers representative of those used in controlling DAP process emissions. Table 41 is a general process description for the dryer, and covers both primary and secondary scrubbers. The primary scrubber may be either a two-stage cyclonic scrubber or a Venturi-cyclone, and the secondary scrubber is limited to a corss-flow packed scrubber. A single efficiency level is requested for each of the two primary scrubbers, while two levels of efficiency are requested for the secondary or tail gas scrubber.

In order to cover the size range encountered for scrubbers used to treat reactor and cooler gas streams as well as those of the dryer, the manufacturers of the scrubbing equipment were asked to provide cost information for a 20,000 DSCFM scrubber (which represents the reactor-drier gas flow on a 1,000 ton/day DAP process), but which is identical in all respects except gas flow capacity to the dryer effluent scrubbers. This should provide costs described for a range of scrubber sizes from which estimates of the cost of treating the reactor-granulator, the dryer, or the cooler vents can be made for most plant sizes.

TABLE 46

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

FLOW, DSCFM	20,000	30,000	48,000	
Effluent Gas Flow				
ACFM	37,000	55,700	89,000	
°F	180	180	180	
SCFM	30,800	46,100	73,800	
Moisture Content, Vol. %	35	35	35	
Effluent Dust Loading				
gr/SCF	0.5	0.5	0.5	
lb/hr, Particulate	133	200	320	
lb/hr, Ammonia	111	167	267	
Cleaned Gas Flow				
ACFM	35,600	53,500	85,500	
°F	164	164	164	
SCFM	31,200	46,800	74,800	
Moisture Content, Vol. %	36	36	36	
Cleaned Gas Dust Loading				
gr/SCF	0.01	0.01	0.01	
lb/hr, Particulate	2.7	4	6.4	
Cleaning Efficiency, % Part.	98	98	98	
Ammonia	99	99	99	
(1) Gas Cleaning Device Cost	48,900*	66,400	92,867	
(2) Auxiliaries Cost	14,050*	21,767	33,900	
(a) Fan(s)		17,350**	30,400**	
(b) Pump(s)		3,850**	5,800**	
(c) Damper(s)		1,450**	2,150**	
(d) Conditioning, Equipment		--	--	
(e) Dust Disposal Equipment		--	--	
(3) Installation Cost	91,700*	102,417	145,733	
(a) Engineering		6,750**	10,500**	
(b) Foundations & Support		19,300**	25,000**	
(c) Ductwork				
(d) Stack		23,500**	33,750**	
(e) Electrical		4,400**	8,000**	
(f) Piping		9,250**	15,350**	
(g) Insulation				
(h) Painting				
(i) Supervision				
(j) Startup		5,750**	9,250**	
(k) Performance Test		2,150**	2,150**	
(l) Other		11,500**	17,000**	
(4) Total Cost	154,650*	190,584	272,500	

* Only two bids obtained for this size.

**Auxiliaries and items of installed cost averaged from two bids. Third bidder did not itemize.

FIGURE 37

CAPITAL COST OF TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

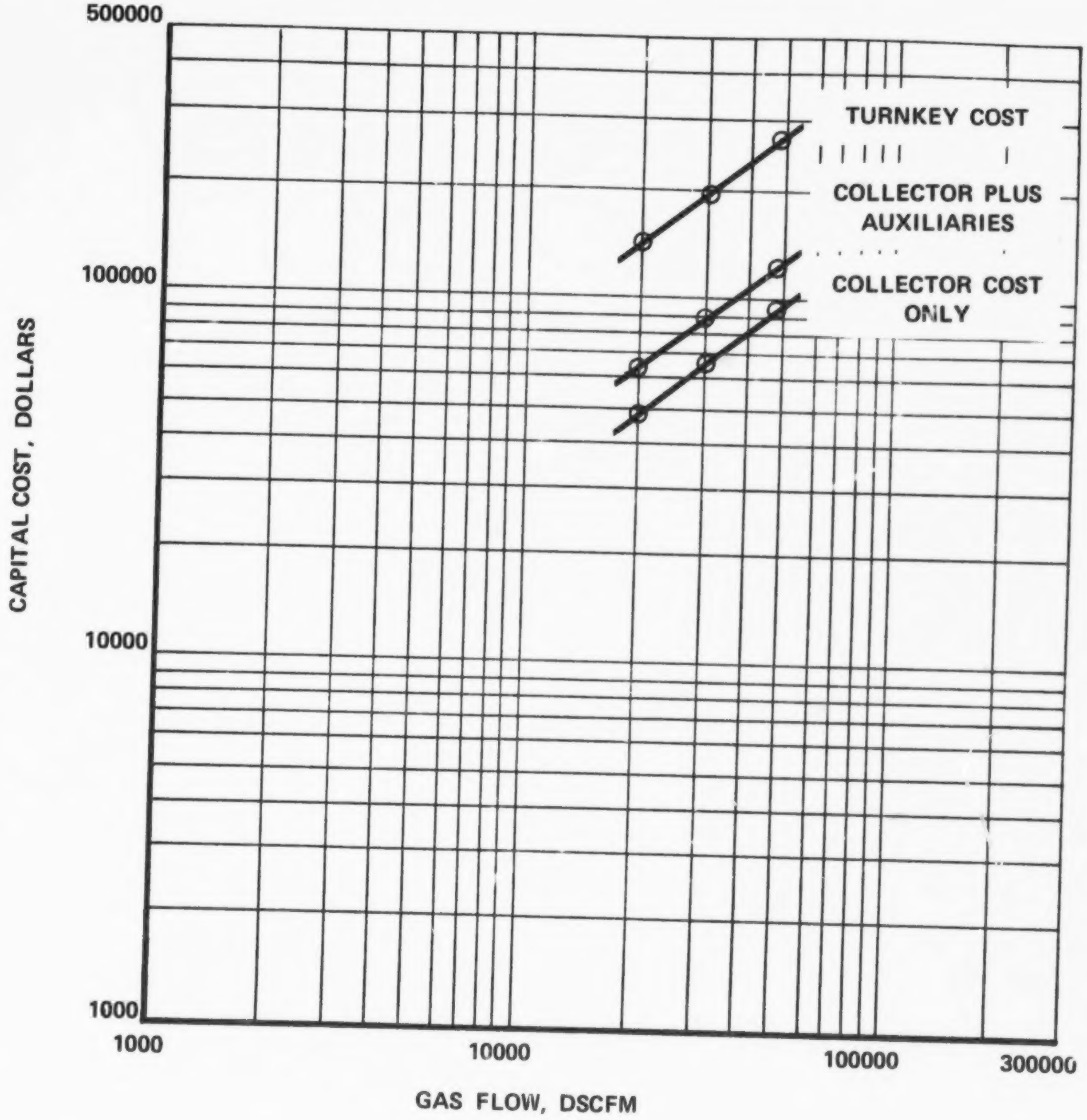


TABLE 47

CONFIDENCE LIMITS FOR CAPITAL COST OF TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

Population Size – 5 Sample Size – 2

Capital Cost = $48,900

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$33,341	$64,459
75	14,444	83,356

Population Size – 5 Sample Size – 3

Capital Cost = $66,400

Conf. Level, %	Lower Limit	Upper Limit
50	$59,121	$73,679
75	52,295	80,504
90	41,793	91,006

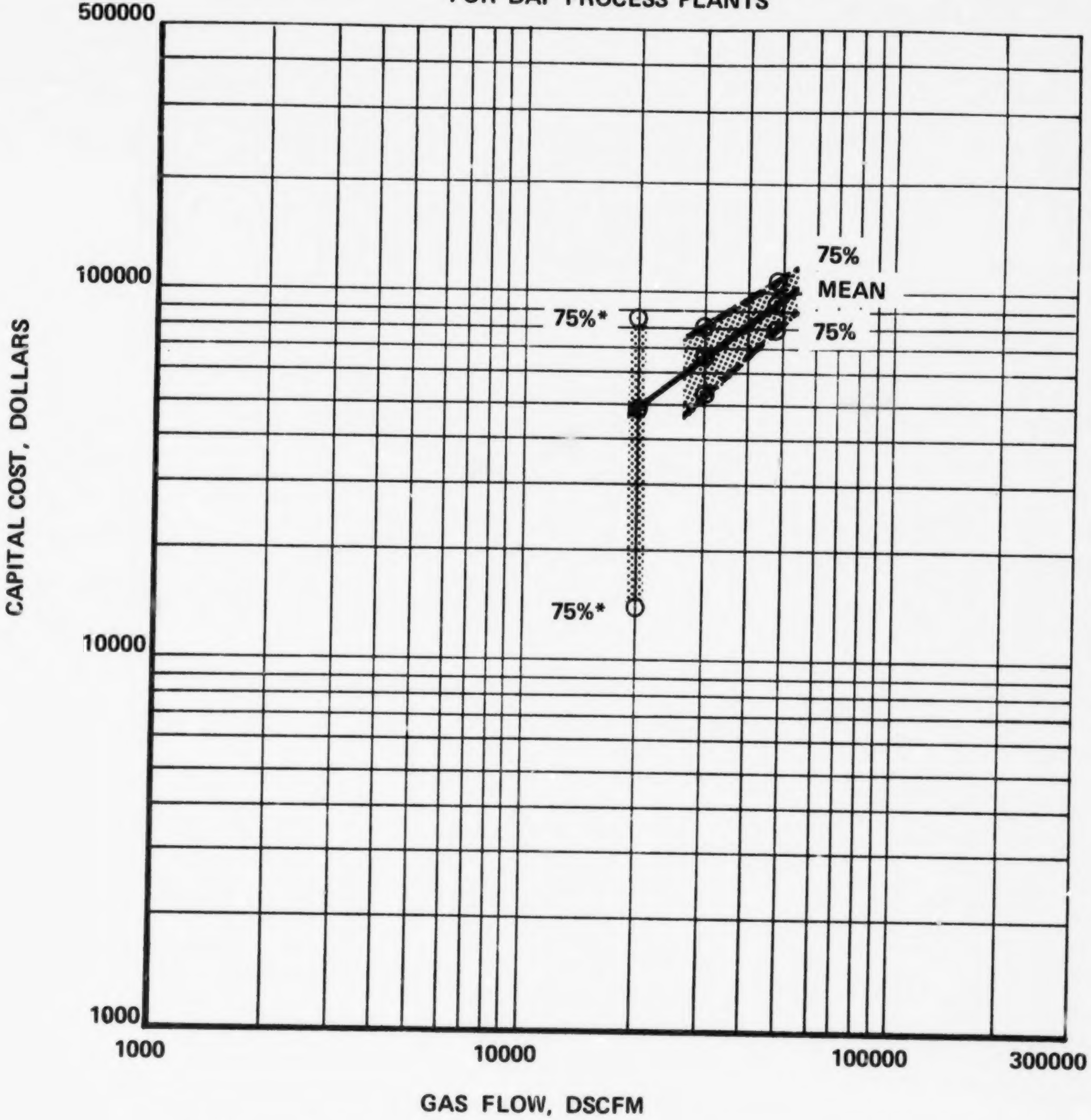
Population Size – 5 Sample Size – 3

Capital Cost = $92,867

Conf. Level, %	Lower Limit	Upper Limit
50	$86,054	$99,679
75	79,667	106,067
90	69,838	115,895

FIGURE 38

CONFIDENCE LIMITS FOR CAPITAL COST OF TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS



*Confidence limit determined from a sample of two bidders. Third bidder did not bid for this size.

TABLE 48

CONFIDENCE LIMITS FOR INSTALLED COST OF TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

Population Size – 5 Sample Size – 2

Installed Cost = $154,650

Conf. Level, %	Installed Cost, Dollars Lower Limit	Upper Limit
50	$128,061	$181,239
75	95,766	213,534
90	32,553	276,747

Population Size – 5 Sample Size – 3

Installed Cost = $190,583

Conf. Level, %	Lower Limit	Upper Limit
50	$176,575	$204,592
75	163,441	217,726
90	143,231	237,936
95	124,542	256,624

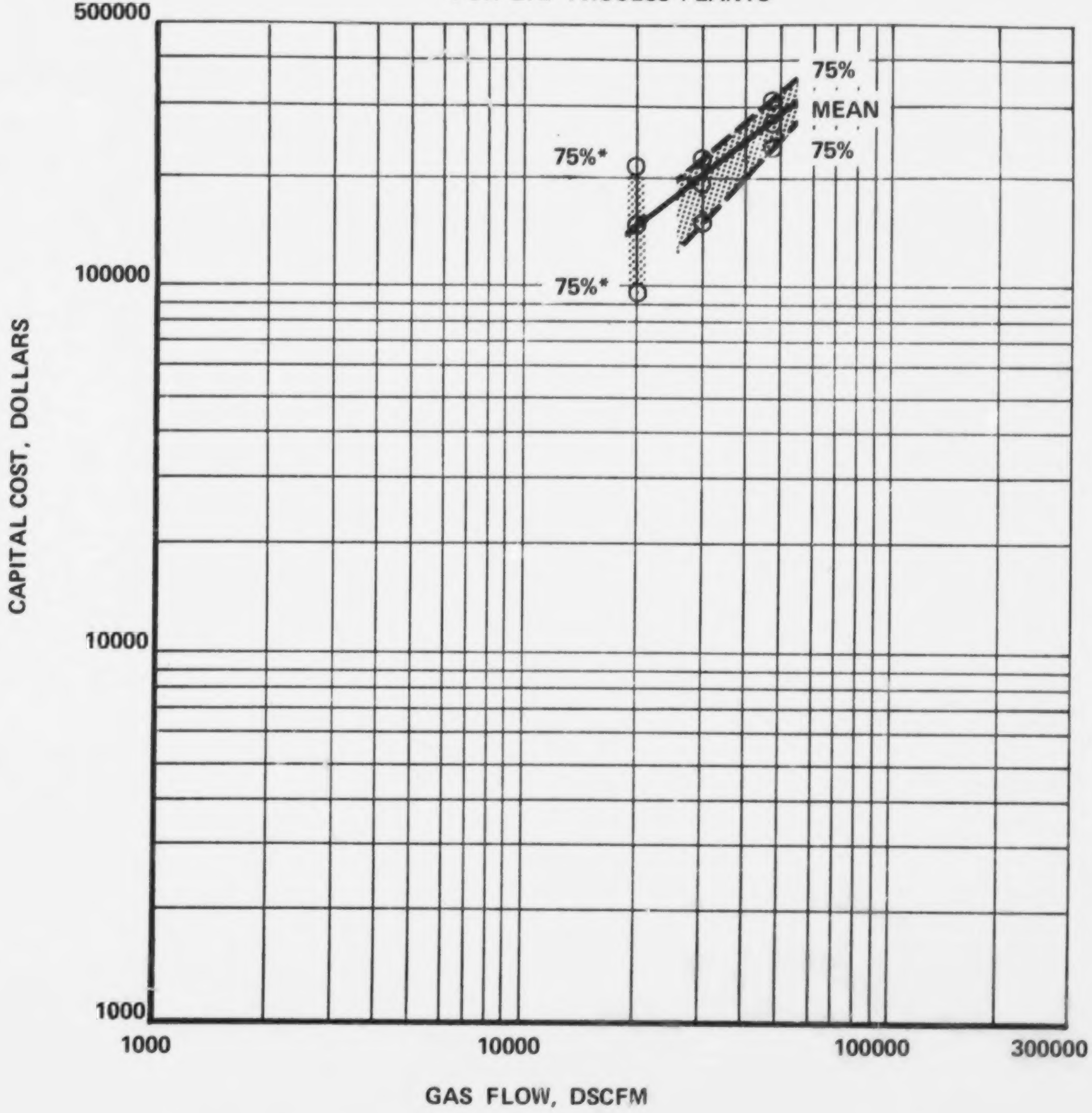
Population Size – 5 Sample Size – 3

Installed Cost = $272,500

Conf. Level, %	Lower Limit	Upper Limit
50	$257,346	$287,654
75	243,137	301,863
90	221,274	323,726
95	201,057	343,943

FIGURE 39

CONFIDENCE LIMITS FOR CAPITAL COST OF INSTALLED TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS



*Confidence limit determined from a sample of two bidders. Third bidder did not bid for this size.

TABLE 49

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

Operating Cost Item	Unit Cost	MODERATE EFFICIENCY			
		20,000 SCFM DG	30,000 SCFM DG	48,000 SCFM DG	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor		$ 3,670*	$ 2,447	$ 2,447	
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		6,825*	6,033	7,450	
Replacement Parts					
Total Replacement Parts		4,229*	3,801	5,355	
Utilities					
Electric Power	$.011/kw-hr	12,030*	15,814	24,920	
Fuel					
Pond Water (Make-up)	$.25/M gal	2,235*	5,259	8,433	
Water (Cooling)	$.05/M gal		--	--	
Chemicals, Specify			--	--	
Total Utilities		14,265	21,073	33,353	
Total Direct Cost		28,989	33,354	48,605	
Annualized Capital Charges		15,465	19,058	27,205	
Total Annual Cost		44,454	52,412	75,810	

* Only two bids obtained for this size.

FIGURE 40

ANNUAL COST OF TWO STAGE CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

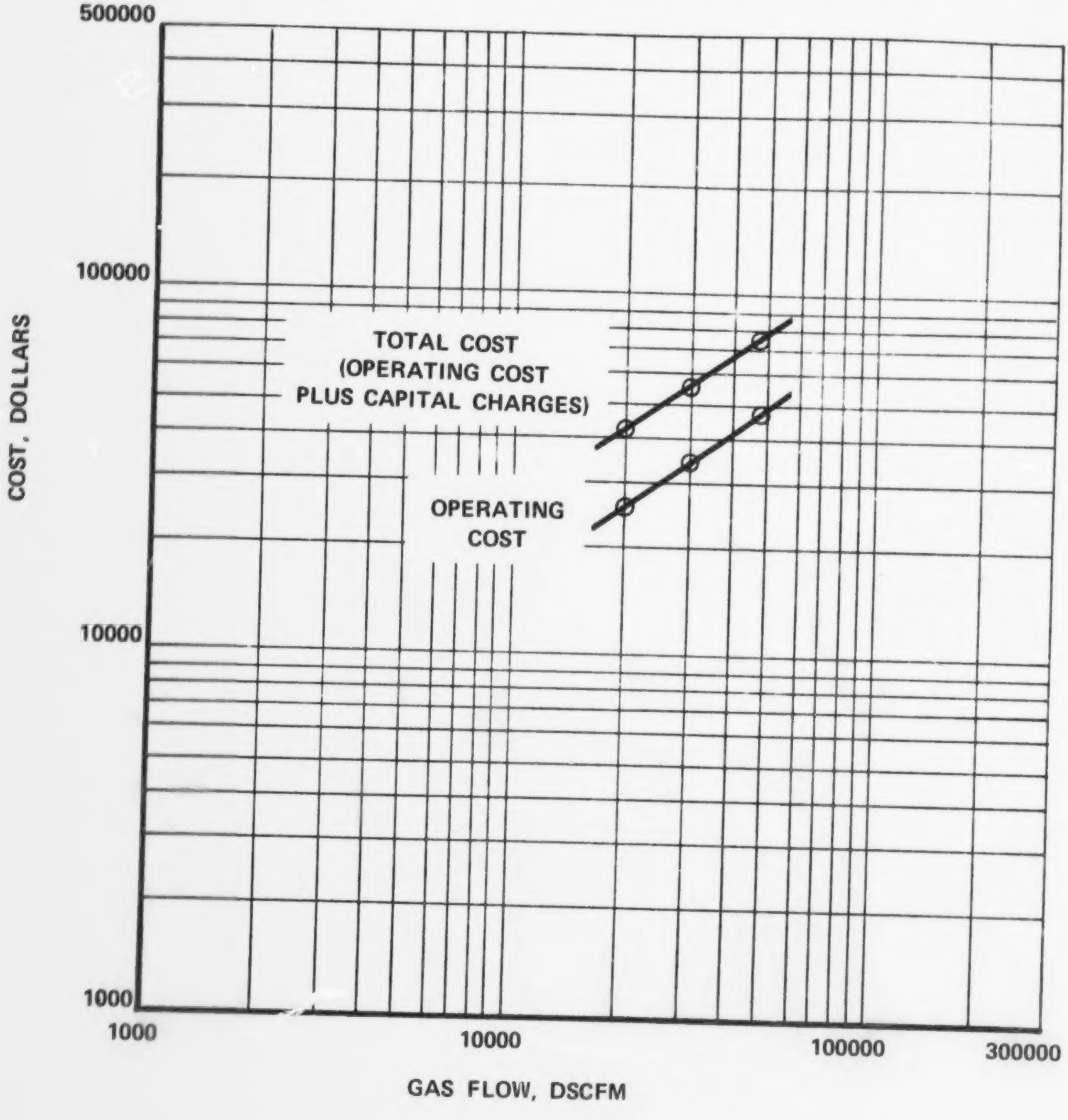


TABLE 50

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR VENTURI CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

FLOW, DSCFM	20,000	30,000	48,000	
Effluent Gas Flow				
ACFM	37,000	55,700	89,000	
°F	180	180	180	
SCFM	30,800	46,100	73,800	
Moisture Content, Vol. %	35	35	35	
Effluent Dust Loading				
gr/SCF	0.5	0.5	0.5	
lb/hr, Particulate	133	200	320	
lb/hr. Ammonia	111	167	267	
Cleaned Gas Flow				
ACFM	35,600	53,500	85,500	
°F	164	164	164	
SCFM	31,200	46,800	74,800	
Moisture Content, Vol. %	36	36	36	
Cleaned Gas Dust Loading				
gr/SCF	0.01	0.01	0.01	
lb/hr, Particulate	2.7	4	6.4	
Cleaning Efficiency, % Part.	98	98	98	
Ammonia	99	99	99	
(1) Gas Cleaning Device Cost	33,250*	44,433	61,467	
(2) Auxiliaries Cost	18,025*	28,500	45,300	
(a) Fan(s)		24,000**	43,100**	
(b) Pump(s)		3,800**	6,700**	
(c) Damper(s)		2,450**	3,150**	
(d) Conditioning, Equipment		--	--	
(e) Dust Disposal Equipment		--	--	
(3) Installation Cost	76,680*	88,833	128,577	
(a) Engineering		6,600**	8,500**	
(b) Foundations & Support		15,000**	21,700**	
(c) Ductwork				
(d) Stack		--	--	
(e) Electrical		13,300**	29,750**	
(f) Piping		5,300**	8,750**	
(g) Insulation		5,800**	8,600**	
(h) Painting				
(i) Supervision		2,100**	2,815**	
(j) Startup				
(k) Performance Test		4,500**	6,250**	
(l) Other		2,150** 14,750**	2,150** 21,600**	
(4) Total Cost	127,955*	161,766	235,344	

* Only two bids obtained for this size.

**Auxiliaries and items of installed cost averaged from two bidders. Third bidder did not itemize.

FIGURE 41

CAPITAL COST OF VENTURI-CYCLONE SCRUBBERS FOR DAP PROCESS PLANTS

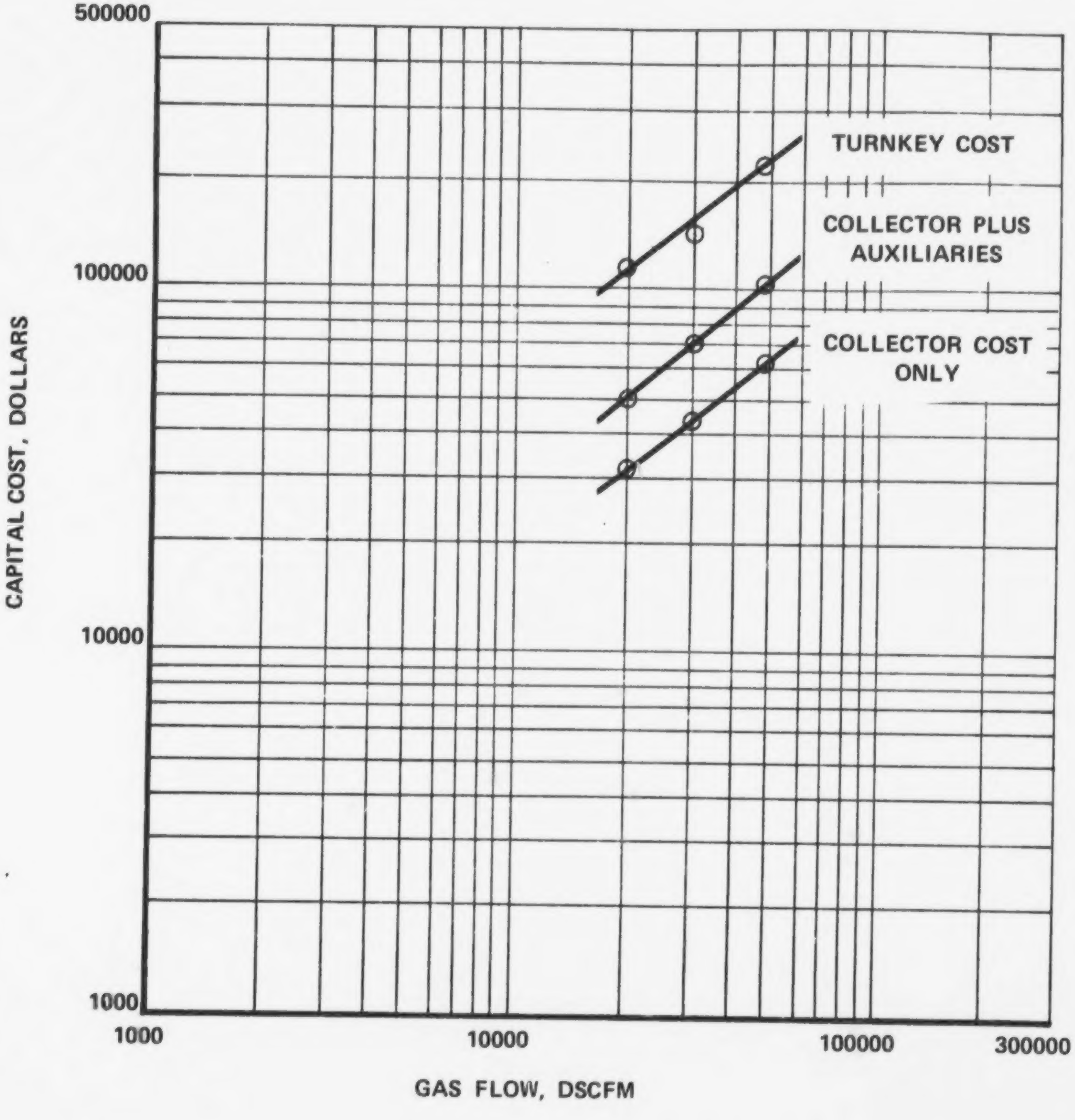


TABLE 51

CONFIDENCE LIMITS FOR CAPITAL COST OF VENTURI-CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

Population Size – 5 Sample Size – 2

Capital Cost = $33,250

Conf. Level, %	Capital Cost, Dollars	
	Lower Limit	Upper Limit
50	$20,284	$46,215
75	4,536	61,963

Population Size – 5 Sample Size – 3

Capital Cost = $44,433

Conf. Level, %	Lower Limit	Upper Limit
50	$38,349	$50,518
75	32,644	56,223
90	23,865	65,001
95	15,748	73,118

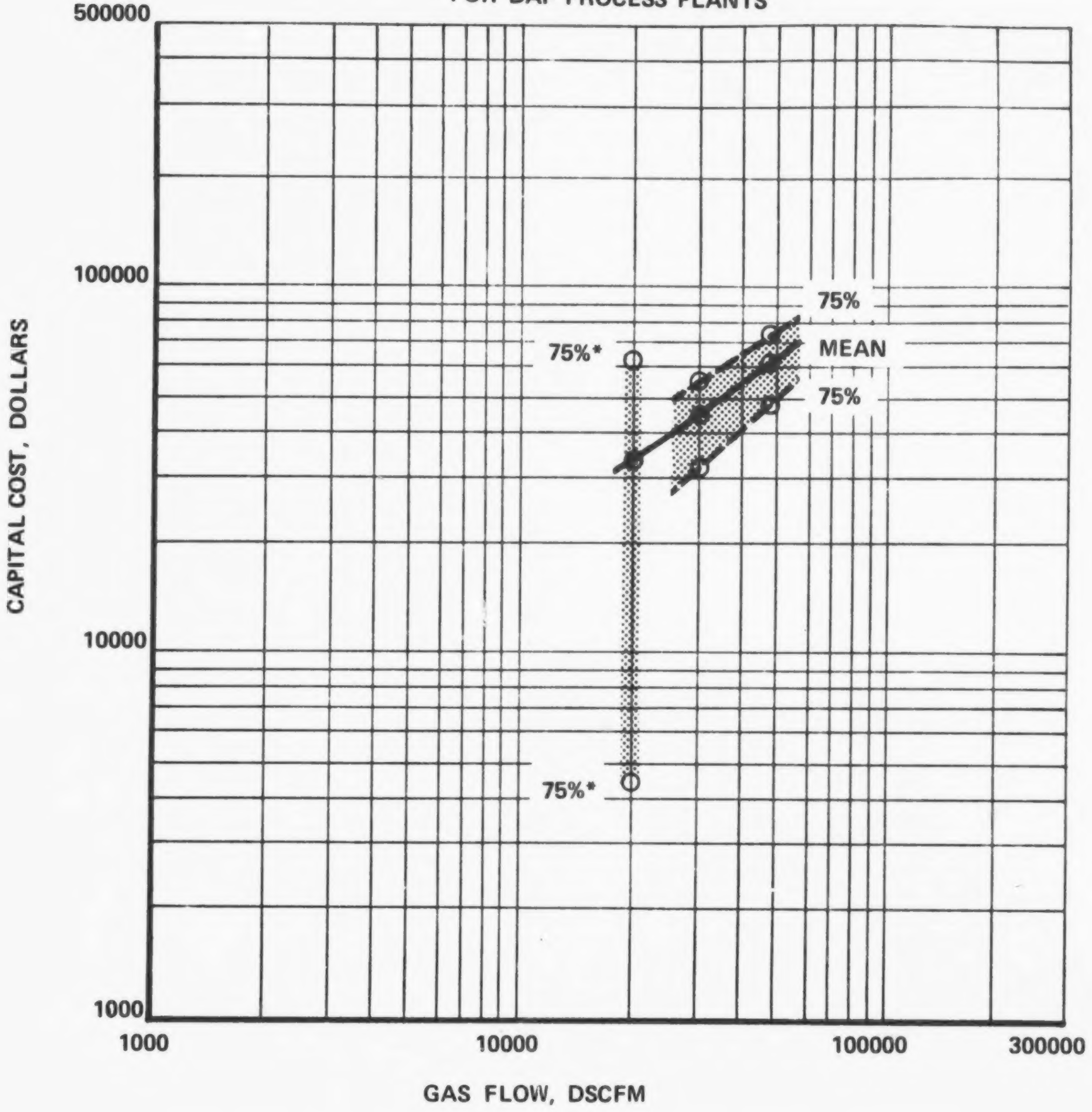
Population Size – 5 Sample Size – 3

Capital Cost = $61,467

Conf. Level, %	Lower Limit	Upper Limit
50	$54,802	$68,131
75	48,553	74,380
90	38,938	83,995
95	30,047	92,886

FIGURE 42

CONFIDENCE LIMITS FOR CAPITAL COST OF VENTURI-CYCLONE SCRUBBERS FOR DAP PROCESS PLANTS



*Confidence limit determined from a sample of two bidders. Third bidder did not bid for this size.

TABLE 52

CONFIDENCE LIMITS FOR INSTALLED COST OF VENTURI-CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

Population Size – 5 Sample Size – 2

Installed Cost = $127,955

Conf. Level, %	Installed Cost, Dollars Lower Limit	Upper Limit
50	$105,472	$150,438
75	78,165	177,745
90	24,715	231,195

Population Size – 5 Sample Size – 3

Installed Cost = $161,767

Conf. Level, %	Lower Limit	Upper Limit
50	$151,791	$171,742
75	142,438	181,095
90	128,046	195,487
95	114,738	208,795

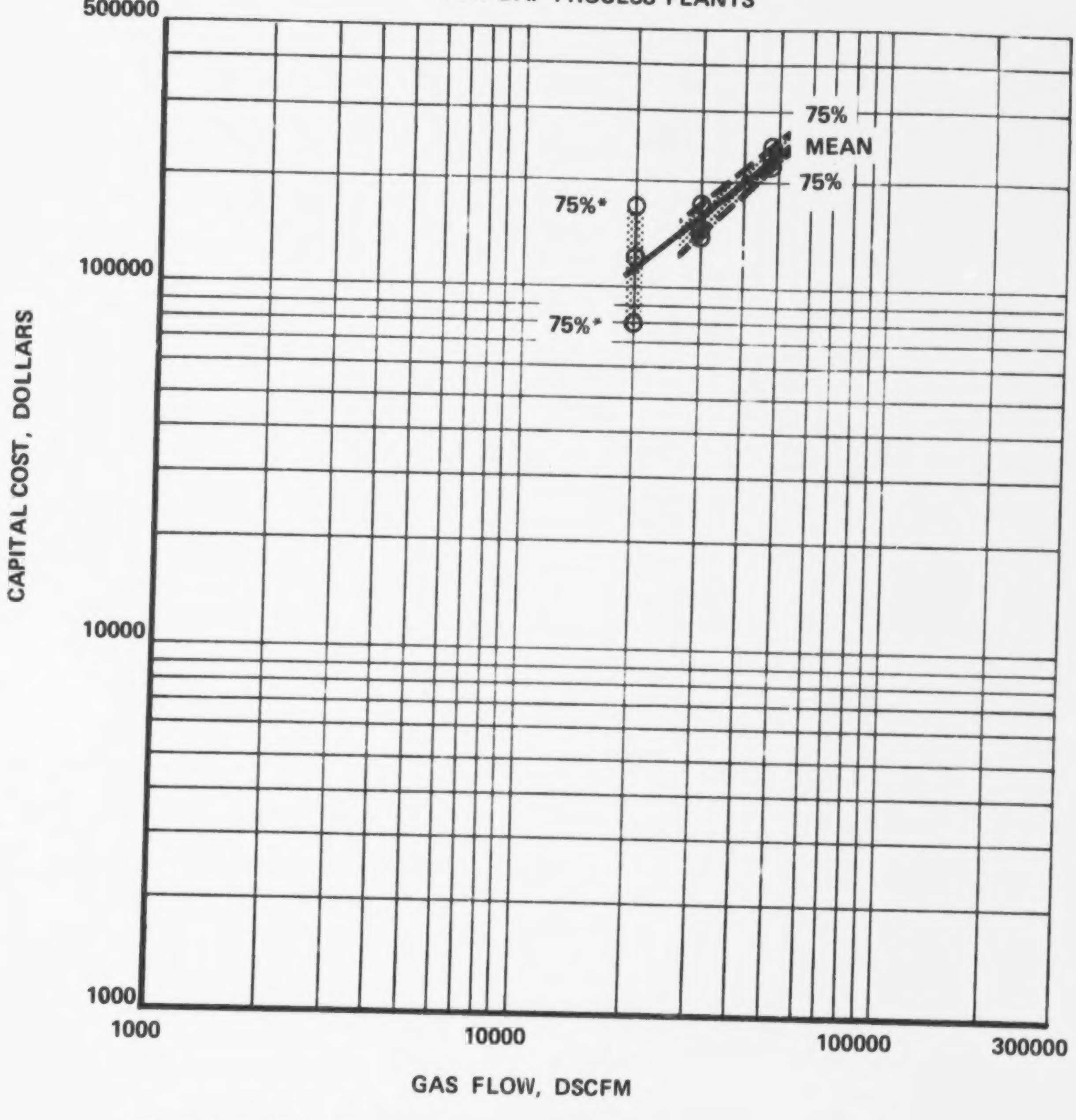
Population Size – 5 Sample Size – 3

Installed Cost = $235,343

Conf. Level, %	Lower Limit	Upper Limit
50	$229,416	$241,271
75	223,859	246,828
90	215,307	255,379
95	207,400	263,287

FIGURE 43

CONFIDENCE LIMITS FOR CAPITAL COST
OF INSTALLED VENTURI-CYCLONE SCRUBBERS
FOR DAP PROCESS PLANTS



*Confidence limit determined from a sample of two bidders. Third bidder did not bid for this size.

TABLE 53

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR VENTURI CYCLONIC SCRUBBERS FOR DAP PROCESS PLANTS

Operating Cost Item	Unit Cost	MODERATE EFFICIENCY			
		20,000 SCFM DG	30,000 SCFM DG	48,000 SCFM DG	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor		$ 2,925*	$ 1,950	$ 1,950	
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		4,650*	4,100	5,166	
Replacement Parts					
Total Replacement Parts		3,932*	3,511	4,828	
Utilities					
Electric Power	$0.011/kw-hr	17,850*	27,645	43,901	
Fuel					
Pond Water (Make-up)	$.25/M gal	1,425***	5,562**	8,950**	
Water (Cooling)	$.05/M gal		--	--	
Chemicals, Specify			--	--	
Total Utilities		18,808*	31,354	49,868	
Total Direct Cost		30,315*	40,915	61,812	
Annualized Capital Charges		12,795*	16,177	23,534	
Total Annual Cost		43,110*	57,092	85,346	

* Only two bids obtained for this size.

**Average of two bidders. Third bidder did not itemize.

***From one bidder only.

FIGURE 44

ANNUAL COST OF VENTURI-CYCLONE SCRUBBERS FOR DAP PROCESS PLANTS

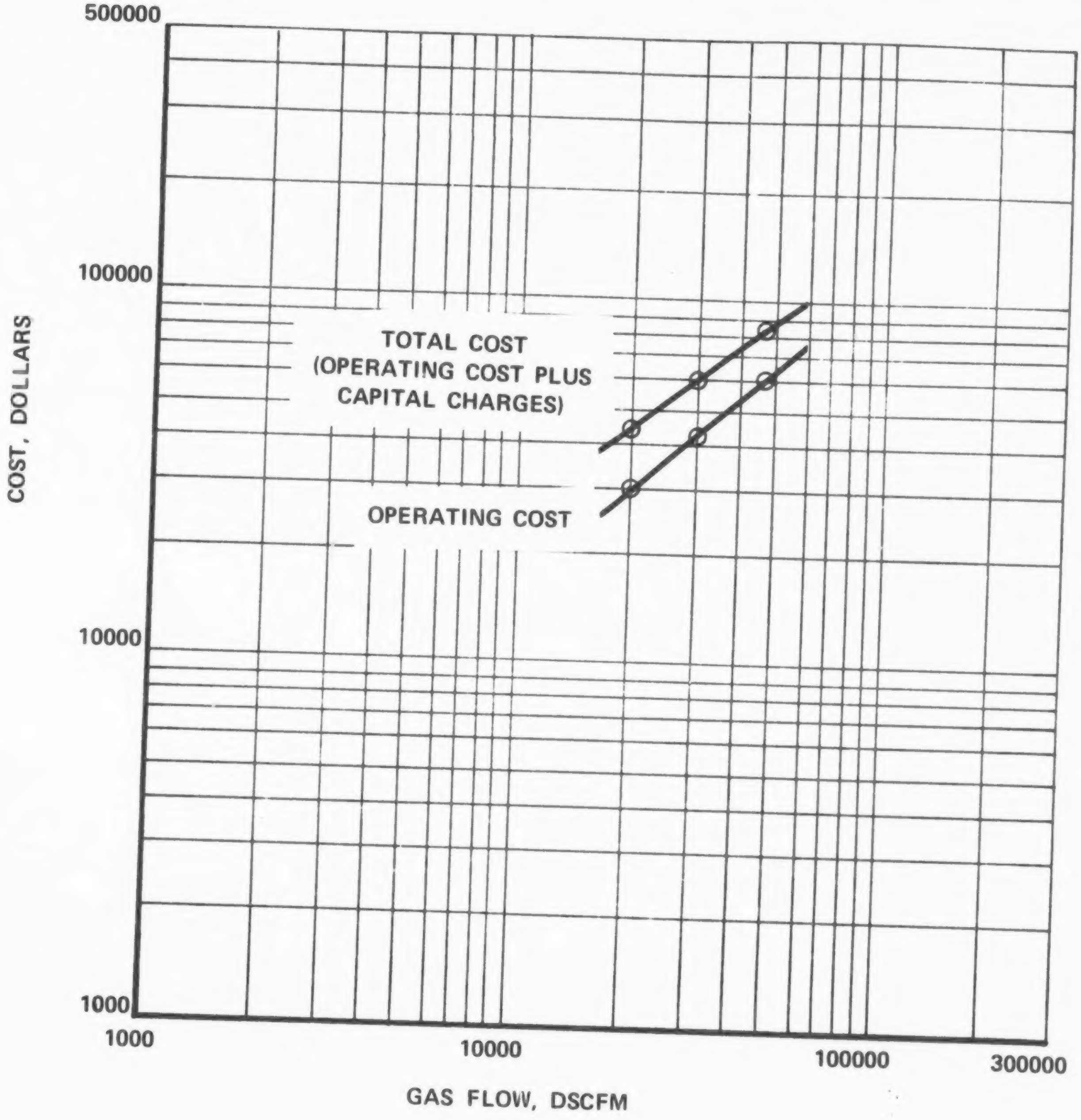


TABLE 54

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS

	MEDIUM EFFICIENCY		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	53,500	85,500	53,500	85,500
°F	164	164	164	164
SCFM	46,800	75,000	46,800	75,000
Moisture Content, Vol. %	36	36	36	36
Contaminant Effluent Loading				
(ppm) Fluorine	15	15	15	15
lb/hr	2.1	3.35	2.1	3.35
Cleaned Gas Flow				
ACFM	33,700	54,000	33,700	54,000
°F	100	100	100	100
SCFM	31,900	51,000	31,900	51,000
Moisture Content, Vol. %	7	7	7	7
Contaminant Cleaned Gas Loading				
(ppm) Fluorine	4.1	4.1	3.25	3.25
lb/hr	0.415	0.67	0.31	0.50
Cleaning Efficiency, %	80	80	85	85
(1) Gas Cleaning Device Cost	27,425	43,200	32,375	47,050
(2) Auxiliaries Cost	13,684	17,470	13,684	17,470
(a) Fan(s)	6,425	9,400	6,425	9,400
(b) Pump(s)	7,089	7,755	7,089	7,755
(c) Damper(s)	170	315	170	315
(d) Conditioning, Equipment	--	--	--	--
(e) Dust Disposal Equipment	--	--	--	--
(3) Installation Cost	38,462	50,358	38,962	50,857
(a) Engineering	2,000	2,000	2,000	2,000
(b) Foundations & Support	3,500*	5,000*	3,500*	5,000*
(c) Ductwork	5,000*	6,200*	5,000*	6,200*
(d) Stack	5,500*	6,000*	5,500*	6,000*
(e) Electrical	5,430*	7,970*	5,430*	7,970*
(f) Piping	10,550*	15,200*	10,550*	15,200*
(g) Insulation	375	500	375	500
(h) Painting	375	450	375	450
(i) Supervision	450	450	450	450
(j) Startup	710	710	710	710
(k) Performance Test	1,475	1,475	1,475	1,475
(l) Other	1,500	2,000	1,500	2,000
(4) Total Cost	79,471	111,028	85,021	115,377

*Items of installed cost. Itemized for materials and labor by one bidder only.

Preceding page blank

TABLE 54 continued

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS

	MEDIUM EFFICIENCY		High Efficiency	
	Small	20,000 DSCFM	Small	20,000 DSCFM
Effluent Gas Flow				
ACFM		36,500		36,500
°F		164		164
SCFM		30,000		30,000
Moisture Content, Vol. %		36		36
Contaminant Effluent Loading				
(ppm) Fluorine		15		15
lb/hr		1.4		1.4
Cleaned Gas Flow				
ACFM		22,600		22,600
°F		100		100
SCFM		21,400		21,400
Moisture Content, Vol. %		7		7
Contaminant Cleaned Gas Loading				
(ppm) Fluorine		4.1		3.25
lb/hr		0.27		0.216
Cleaning Efficiency, %		80		85
(1) Gas Cleaning Device Cost		21,025		23,525
(2) Auxiliaries Cost		9,502		9,502
(a) Fan(s)		4,350		4,350
(b) Pump(s)		5,000		5,000
(c) Damper(s)		152		152
(d) Conditioning Equipment		--		--
(e) Dust Disposal Equipment		--		--
(3) Installation Cost		32,540		33,042
(a) Engineering		2,000		2,000
(b) Foundations & Support		3,000*		3,000*
(c) Ductwork		4,600*		4,600*
(d) Stack		3,000*		3,000*
(e) Electrical		4,040*		4,040*
(f) Piping		6,200*		6,250*
(g) Insulation		325		325
(h) Painting		325		325
(i) Supervision		450		450
(j) Startup		710		710
(k) Performance Test		1,475		1,475
(l) Other		1,250		1,250
(4) Total Cost		63,067		66,069

*Items of installed cost. Itemized for materials and labor by one bidder only.

FIGURE 45

CAPITAL COST FOR PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS

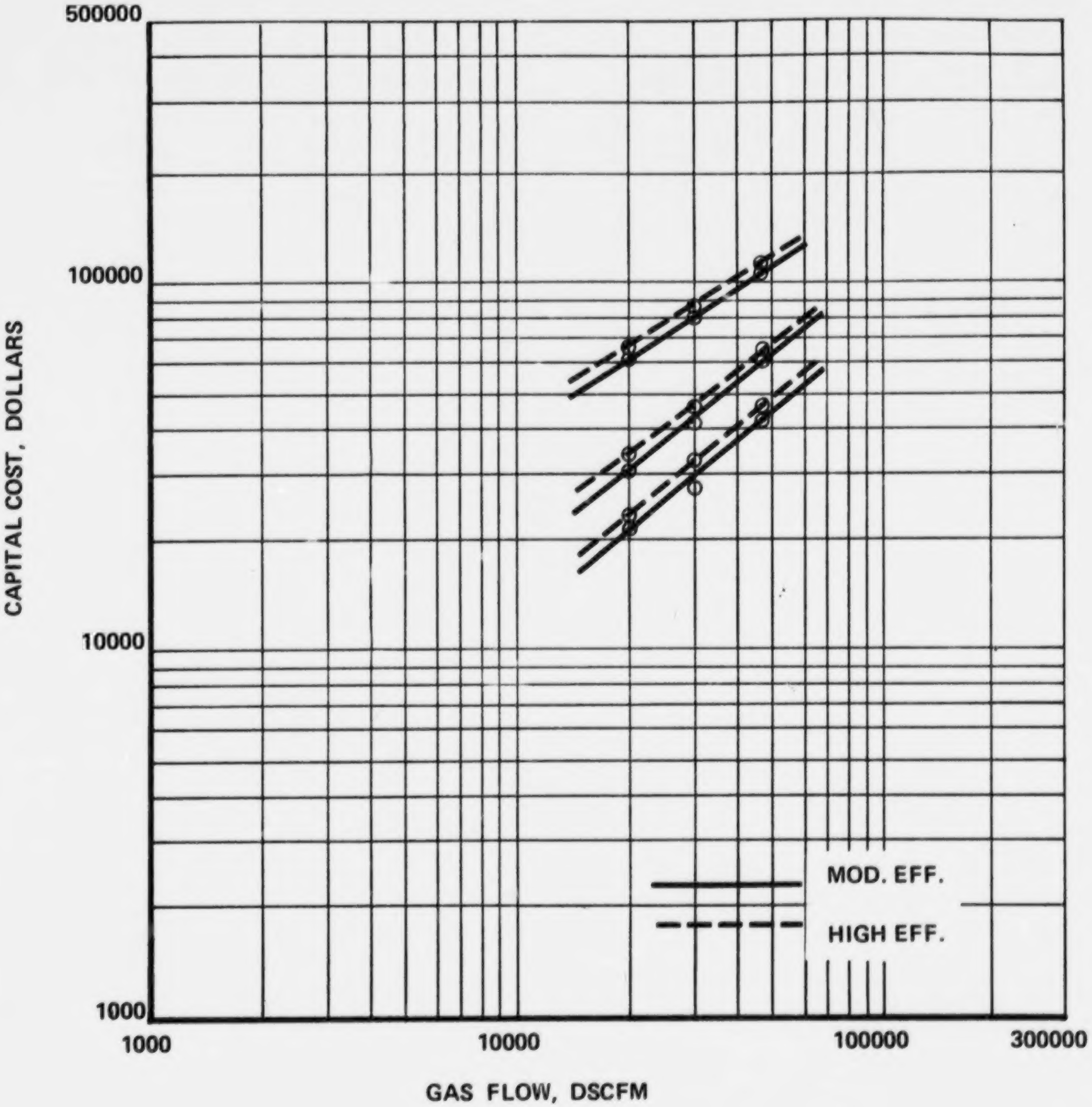


TABLE 55

CONFIDENCE LIMITS FOR CAPITAL COST OF PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS

Population Size – 5 Sample Size – 2

Capital Cost = $21,025

Conf. Level %	Capital Cost, Dollars Lower Limit	Capital Cost, Dollars Upper Limit
50	20,001	22,049
75	18,757	23,293
90	16,322	25,729
95	13,691	28,359

Population Size – 5 Sample Size – 2

Capital Cost = $27,425

Conf. Level %	Capital Cost, Dollars Lower Limit	Capital Cost, Dollars Upper Limit
50	22,603	32,247
75	16,747	38,103
90	5,283	49,566

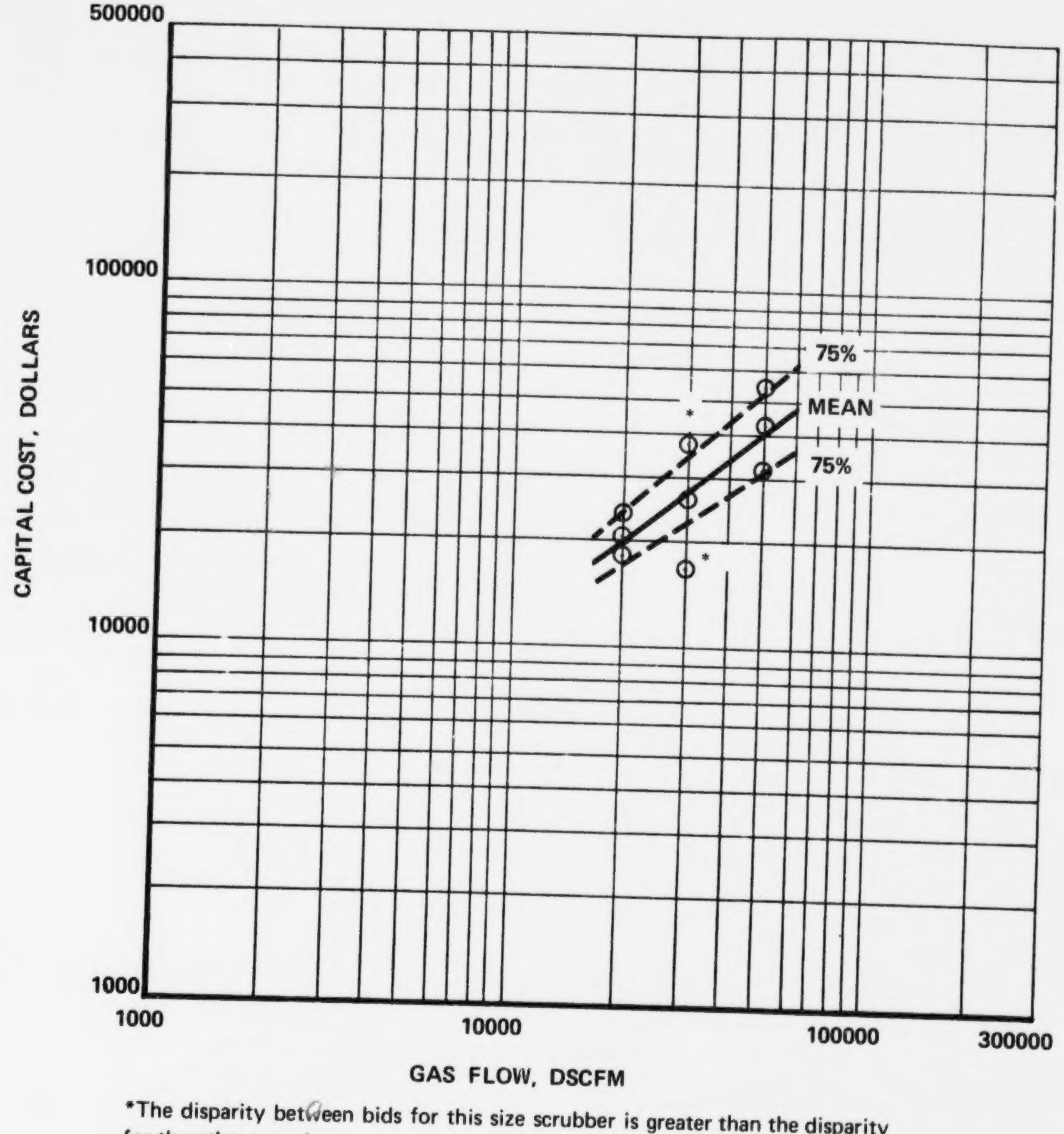
Population Size – 5 Sample Size – 2

Capital Cost = $43,200

Conf. Level %	Capital Cost, Dollars Lower Limit	Capital Cost, Dollars Upper Limit
50	38,004	48,396
75	31,692	54,708
90	19,338	67,062
95	5,995	80,405

FIGURE 46

CONFIDENCE LIMITS FOR CAPITAL COST
OF PACKED CROSS-FLOW SCRUBBERS
FOR DAP PROCESS PLANTS



*The disparity between bids for this size scrubber is greater than the disparity for the other two sizes.

TABLE 56

CONFIDENCE LIMITS FOR INSTALLED COST OF PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS

Population Size – 5 Sample Size – 2

Installed Cost = $63,070

Conf. Level %	Lower Limit	Upper Limit
50	61,596	64,544
75	59,806	66,334
90	56,302	69,839
95	52,517	73,623

Population Size – 5 Sample Size – 2

Installed Cost = $79,572

Conf. Level %	Lower Limit	Upper Limit
50	71,151	87,992
75	60,923	98,220
90	40,904	118,239
95	19,281	139,861

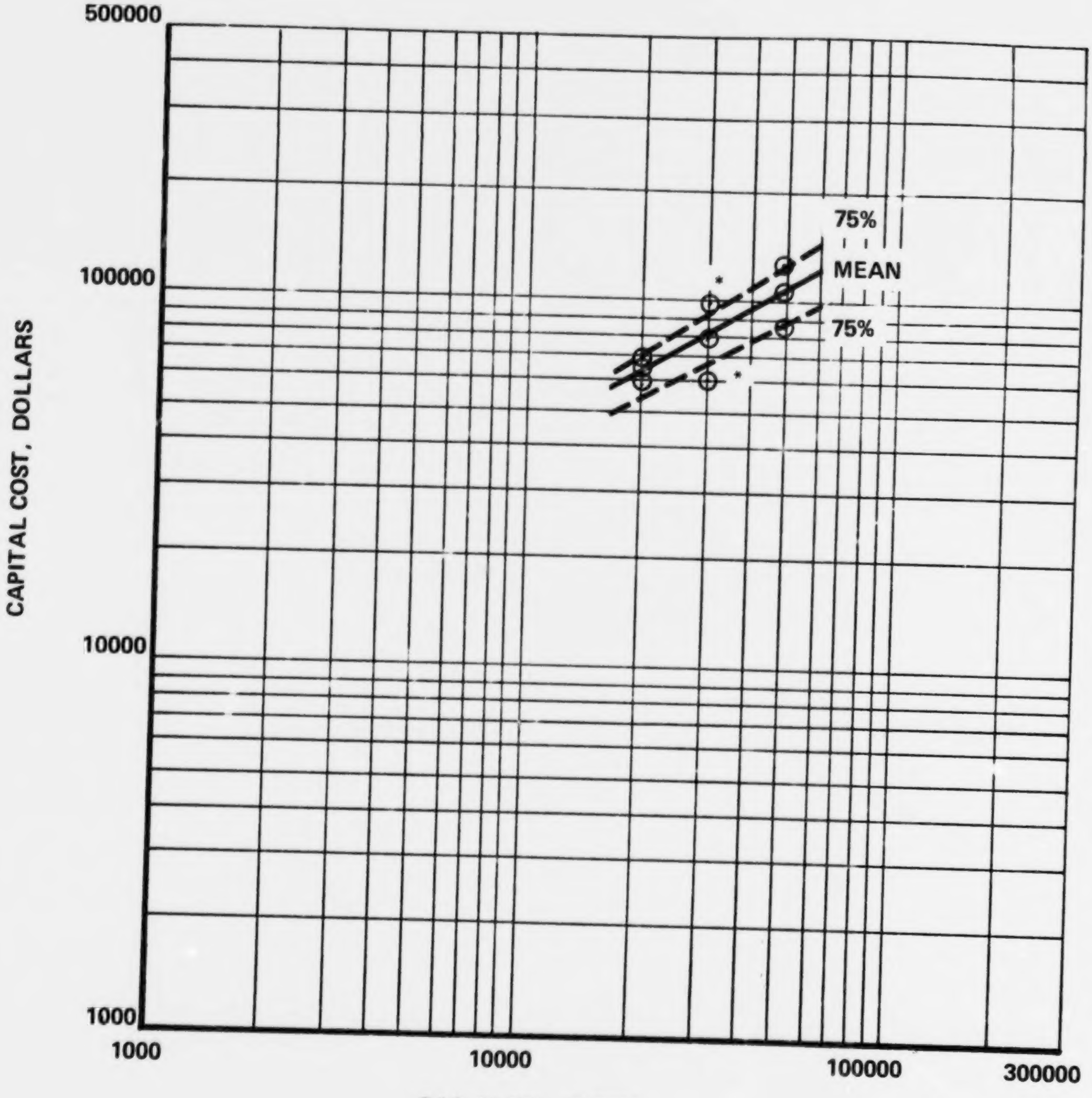
Population Size – 5 Sample Size – 2

Installed Cost = $111,028

Conf. Level	Lower Limit	Upper Limit
50	100,552	121,503
75	87,829	134,226
90	62,925	159,130
95	36,027	186,028

FIGURE 47

CONFIDENCE LIMITS FOR INSTALLED COST OF PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS



* The disparity between bids for this size scrubber is greater than the disparity for the other two sizes.

TABLE 57

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR)
FOR PACKED CROSS-FLOW SCRUBBERS
FOR DAP PROCESS PLANTS

Operating Cost Item	Unit Cost	MEDIUM EFFICIENCY		High Efficiency	
		30,000 DSCFM	48,000 DSCFM	30,000 DSCFM	48,000 DSCFM
Operating Factor, Hr/Year		8,000	8,000	8,000	8,000
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor		1,738	3,613	1,738	3,613
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		1,875	2,888	1,875	2,888
Replacement Parts					
Total Replacement Parts		200	250	200	250
Utilities					
Electric Power	$.011/kw-hr	4,787	7,150	4,990	7,287
Fuel					
Pond Water (Make-up)	$.25/M gal	34	55	35	57
Water (Cooling)					
Chemicals, Specify	$.05/M gal				
Total Utilities		4,821	7,205	5,025	7,344
Total Direct Cost		8,624	13,955	8,828	14,095
Annualized Capital Charges		7,957	11,103	8,502	11,538
Total Annual Cost		16,581	25,058	17,330	25,633

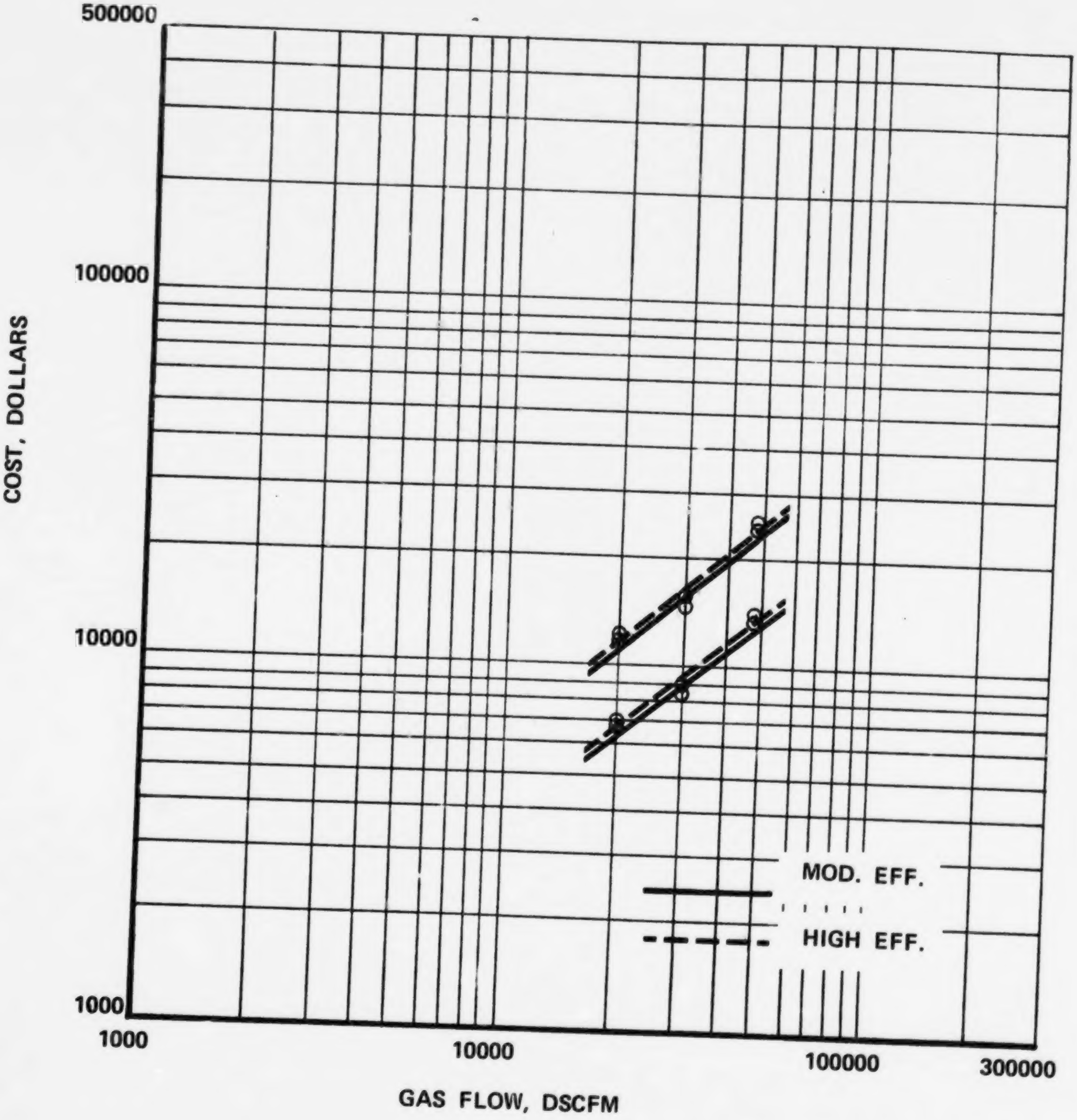
TABLE 57 continued

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS

Operating Cost Item	Unit Cost	MEDIUM EFFICIENCY		High Efficiency	
			.20,000 DSCFM		20,000 DSCFM
Operating Factor, Hr/Year			8,000		8,000
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor			1,738		1,738
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance			1,838		1,838
Replacement Parts					
Total Replacement Parts			150		150
Utilities					
Electric Power	$.011/kw-hr		3,051		3,183
Fuel					
Pond Water (Make-up)	$.025/M gal		23		24
Water (Cooling)	$.05/M gal				
Chemicals, Specify					
Total Utilities			3,074		3,207
Total Direct Cost			6,800		6,933
Annualized Capital Charges			6,307		6,609
Total Annual Cost			13,107		13,540

FIGURE 48

ANNUAL COST OF PACKED CROSS-FLOW SCRUBBERS FOR DAP PROCESS PLANTS



GTSP STORAGE

The reactions initiated in the GTSP reactor are completed over a period of several weeks after acidulation.(6) During this time the granular product must be stored under roof to protect it from the weather. The storage building must be well ventilated to provide a safe environment for employees engaged in loading, unloading and moving the bulk material, and the ventilating air stream must be treated to remove dust and fluorine (or SiF_4) before discharge to the atmosphere.

Typically the storage is in gable-roofed shed buildings which are ventilated by natural draft if no air pollution control equipment is used. Such buildings are designed for high ventilation rates, typically from 100,000 to 1,000,000 SCFM, with thermal density differences as the driving force. The addition of scrubbing equipment for air pollution control provides a strong incentive for the reduction of the ventilation rate to the minimum consistent with good working conditions. This section discusses the problem in terms of treating a reduced flow of about 200,000 ACFM.

Process Description

The screened granular triplesuperphosphate is ordinarily discharged from the last screening operation onto a rubber conveyor belt which carries it into the storage building. Here it is discharged into bins or piles. The product may be moved into position in a larger bulk storage pile by use of portable elevators, overhead cross transfer belts, front-end loaders, or by other relocatable solids handling equipment.

Any scheme for moving the solids inside the storage building must take into account that all the GTSP should be held long enough to stabilize the composition so no fluorine evolution will take place after the product is shipped. Any additional storage time is wasteful of space. The logistics of moving 500 or 1,000 ton/day solid material through a 30-day retention pile requires a well planned solids handling program.

Usually the buildings are designed for holding a pyramid-shaped pile of GTSP considerably larger than that required for a 30-day retention. This allows for some inefficiency in use of the space to minimize handling.

The evolution of fluorine as SiF_4 occurs slowly, and at a rate which decreases with storage time. If the entire storage building is treated as a unit, the evolution rate has been quoted as approximately 0.2 to 0.6 lb F day/ton of

GTSP.(6) Thus, a storage building designed for a throughput of 500 ton/day (P_2O_5 content) might be expected to evolve 100 to 300 lb/day of SiF_4.

The size of the building determines the ventilation rate required to produce good working conditions inside. For a 500 ton/day plant, the average storage level should be around 15,000 ton, and the building should have an interior volume of perhaps three times the average value. At 30 lb/ft^3, the volume should be on the order of

$$\frac{15{,}000 \times 3 \times 2000}{0.564 \times 40} = 4{,}000{,}000 \text{ ft}^3$$

A building of the dimensions shown in Figure 28 would contain approximately this volume.

Ventilation Requirements

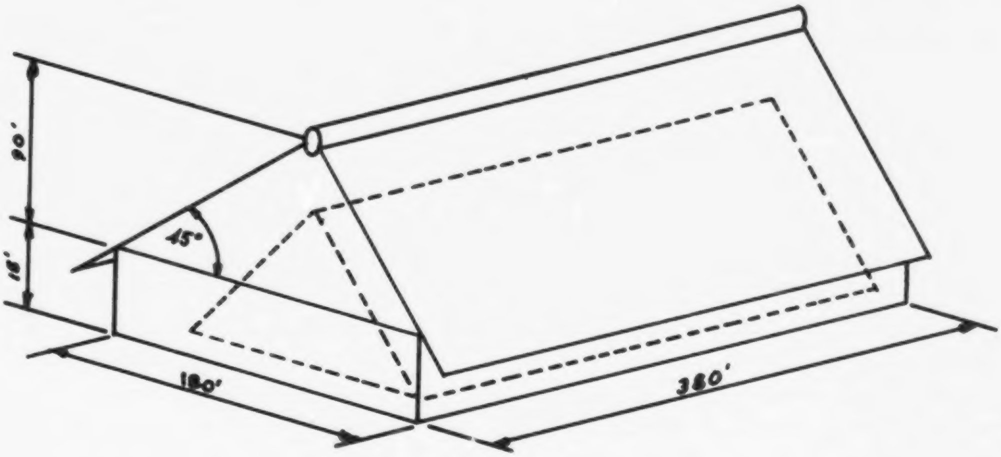
Ventilation for worker safety in storage buildings is ordinarily expressed in terms of the number of "air changes per hour", based on total building volume. Typical values for warehouses range from 1 to 15 air changes per hour,(17) or one complete turnover of the air inside the building in 60 minutes for the lower rate and 4 minutes for the higher one. For the 4,000,000 cubic foot storage building this amounts to a minimum rate of 67,000 ACFM and a maximum of 1,000,000 ACFM.

The emissions of SiF_4 require a higher than minimum rate of ventilation,(6) and three air changes per hour is a reasonable value. This corresponds to 200,000 ACFM for a hypothetical 500 ton/day GTSP plant. Careful selection of the ventilation rate is important because the cost of gas cleaning equipment varies in direct relation to the gas flow, and oversizing the ventilation system will increase the cost of purchasing and operating the gas cleaning equipment.

Design of the ventilation system in such a way as to produce acceptable air quality in all parts of the storage building is complicated by several factors:

(1) The dust release is primarily local and associated with handling of the stored product.

(2) The fluorine release is slow and decreases with time.



90'
18'
45°
180'
380'

FIGURE 49

SKETCH OF GTSP STORAGE BUILDING WITH 4,000,000 FT^3 INTERIOR

The building ventilation system should be designed to provide for clean air induction in a controlled pattern as well as adequate exhaust of contaminated air.

Pollution Control Considerations

The dust generated by solids handling and the fluorine evolved by curing of the TSP product pose potential pollution problems.(8) These can be treated separately by use of a fabric collector for particulate removal followed by a cyclonic scrubber or packed scrubber for gas absorption, or by a scrubber which serves both functions. The latter case is the only one considered here, but combination systems are practical and have been used.

The particulate matter generated by mechanical handling of the granular TSP is relatively fine with the coarser particles separated out of the ventilating air stream by gravity. Particle size is predominantly above one micron, however, any moderate scrubber pressure drop levels (less than 10" w.c.) are adequate for collection. Typical dust loadings on the order of 0.1 gr/ACF or

$$\frac{0.1 \times 200{,}000 \times 60}{7{,}000} = 170 \text{ lb/hr}$$

are encountered, but the rate of dust evolution is likely to be very variable and very low levels of dust generation are likely when no materials handling operations are being carried out. Peak emission rates, on the other hand, may reach levels of 1 gr/ACF for short periods, and require efficiencies on the order of 99% to avoid visible discharge and violation of particulate emission limits.

The gas absorption requirements are relatively straightforward, and may be met by use of a variety of absorption devices. The emission from a GTSP storage building for a 500 ton/day plant will be in the range of 0.5 lb/ton of P_2O_5,(6) which is approximately 250 lb/day of fluorine. The present Florida law limits the total emission to 0.05 lb/ton P_2O_5, or approximately 25 lb/day. This requires 90% scrubbing efficiency and an outlet concentration of about

$$\frac{\frac{25}{19} \times \frac{379}{60} \times \frac{1}{24}}{200{,}000} \times 1{,}000{,}000 = 1.73 \text{ ppm},$$

equivalent to 0.04 mg/SCF as F. This efficiency level demands low temperature pond water in large quantity for scrubbing. Reference to Figure 8 and Table 13 indicates that equilibrium levels of fluorine are low enough to meet the required outlet concentration at temperatures on the order of 80° F at almost any pond water fluorine content. At 100° F, a low concentration of H_2SiF_6 (below 1.0 wt.%) would be required, and at 120° F the specification could not be met.

Field experience has shown that the header ducts from storage building should be sprayed to minimize solids build-up. Further, the spraying converts the SiF_4 to H_2SiF_6 and SiO_2 which then becomes conditioned for scrubbing, minimizing space otherwise lost in the scrubber shell.

With 80° F pond water of normal fluorine content, the absorption requirement is not severe. For this reason, a variety of scrubbers can perform satisfactorily in this service, and minimization of energy input is more important than maximization of gas-liquid contact. The following scrubber types are acceptable on this basis:

(1) Cyclonic

(2) Spray towers

(3) Cross flow packed

The use of a conventional cyclonic scrubbing stage is relatively common, and the single specification for this service is based on this combination. Figure is a photograph of a GTSP storage building, showing two cyclonic scrubbers.

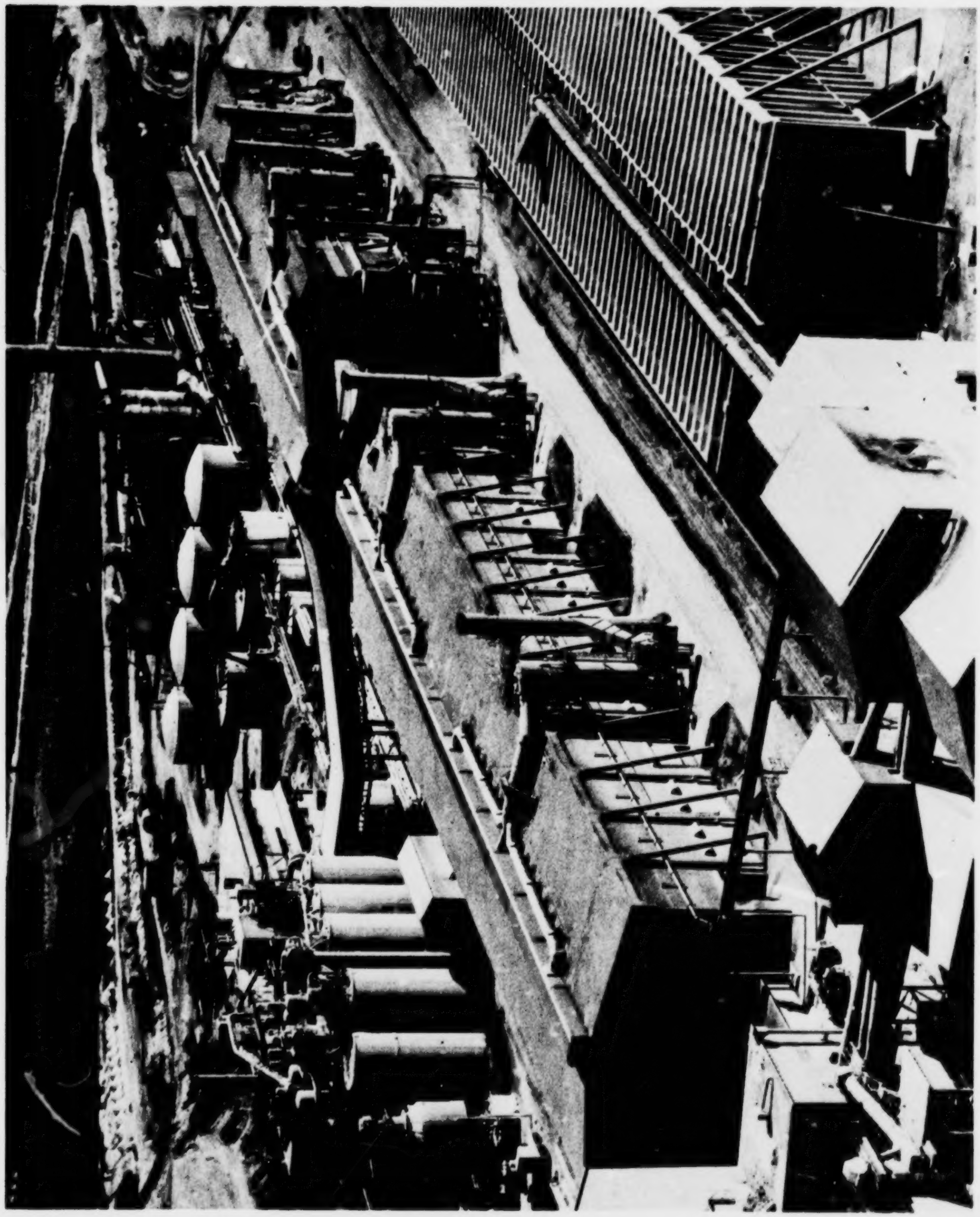


FIGURE 50

GTSP STORAGE BUILDING
WITH CYCLONIC SCRUBBERS

TABLE 58

PROCESS DESCRIPTION FOR GTSP STORAGE BUILDING SCRUBBER

This specification describes a scrubber to remove particulate matter and gaseous fluorides from the ventilating air discharged from a granular triplesuperphosphate fertilizer storage building. The scrubber shall consist of a cyclonic scrubber for absorption of gaseous SiF_4 or SiO_2, particulate collection and de-entrainment of scrubbing liquor.

The scrubbing system is to include a fan to overcome the pressure drop across the scrubber and associated ductwork, a collection plenum system equipped with internal water sprays, which is to connect roof ventilators along the length of the storage building, balancing dampers, scrubber inlet ductwork, a 180 ft. discharge stack, and a liquor return sump. The building dimensions are approximately 180 ft. wide x 380 ft. long x 105 ft. at the ridge line.

Pond water is to be used as the scrubbing medium in both stages. The scrubbing system will discharge into a sump to be provided as a part of the scrubbing system. Pumps for supplying the pond water will be by others, but the scrubber contractor is to supply the pond water return pumps as a part of his system.

Pond water characteristics are as follows:

	Design	*Min.*	*Max.*
Pond Water, pH	2.0	1.2	2.2
Temp., ^{o}F	80	55	88
SO_4, *wt.%*	0.15	–	–
P_2O_5, *wt.%*	0.1	–	–
H_2SiF_6, *wt.%*	0.63	0.25	1.0
Fluorine, wt.%	0.5	0.2	0.8

The scrubber is required to produce the specified performance when operating with the "design" pond water conditions. The scrubber manufacturer shall specify the required water circulation rate through both the duct section and the Cyclonic scrubber.

Materials of construction shall be limited to the following:

PVC
Rubber
FRP (Dynel lined)

Minimal metal parts shall be exposed to process gas or pond water.

Preceding page blank

TABLE 59

OPERATING CONDITIONS FOR SCRUBBER SPECIFICATION FOR GTSP STORAGE BUILDING VENT

Plant Capacity, ton/day P_2O_5	500
Building Dimensions, ft.	
Width	180
Length	380
Height at Ridge	105
Building Volume, ft^3	4,000,000
Ventilation Rate, ACFM	200,000
Gas to Scrubber	
Flow, ACFM	200,000
Temperature, °F	80
Moisture Content, Vol.%	1.6
Flow, DSCFM	196,800
Gaseous Fluoride (SiF_4) Loading, ppm	17.3
Gaseous Fluoride (SiF_4) Loading, lb/hr	10.4
Particulate Solid, gr/ACF	0.10
Particulate Solids, lb/hr	170
Gas from Scrubber	
Flow, ACFM	202,000
Flow, DSCFM	196,800
Temp., °F	68
Moisture Content, Vol.%	2.4
Gaseous Fluoride Loading, ppm	1.70
Gaseous Fluoride Loading, lb/hr	1.0
Fluoride Efficiency, Wt.%	90.4
Particulate Solids Loading, gr/ACF	0.01
Particulate Solids Loading, lb/hr	1.7
Particulate Efficiency, Wt.%	90
Estimated Fluorine in Equilibrium with Pond Water, ppm	0.7
Estimated NTU required	1.8

TABLE 60

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CYCLONIC SCRUBBERS FOR GTSP STORAGE VENTS

	LA Process Wt.		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM				200,000
°F				80
SCFM				196,000
Moisture Content, Vol. %				1.6
Effluent Concentration Loading				
(ppm), Fluoride				17.3
lb/hr, Particulate				170
Cleaned Gas Flow				
ACFM				202,000
°F				68
SCFM				196,800
Moisture Content, Vol. %				2.4
Cleaned Gas Dust Loading				
(ppm), Fluoride				1.70
lb/hr, Particulate				17
Cleaning Efficiency, %				90
(1) Gas Cleaning Device Cost				101,000
(2) Auxiliaries Cost				109,250
(a) Fan(s)				57,700*
(b) Pump(s)				9,975*
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				333,480
(a) Engineering				13,900*
(b) Foundations & Support				21,150*
(c) Ductwork				
(d) Stack			(48,000)**	(312,000)**
(e) Electrical				20,100*
(f) Piping				22,000*
(g) Insulation				
(h) Painting				3,570*
(i) Supervision				10,500*
(j) Startup				5,250*
(k) Performance Test				28,750*
(l) Other				
(4) Total Cost				543,730

* Average of two bidders. Third bidder did not itemize.

**Lower price from one bidder for one stack to handle 200,000 SCFM. Higher price from another bidder for four stacks each handling 50,000 SCFM.

TABLE 61

CONFIDENCE LIMITS FOR CAPITAL COST OF CYCLONIC SCRUBBERS FOR GTSP STORAGE

Population Size – 5

Sample Size – 3

Capital Cost = $101,000

Conf. Level, %	Capital Cost, Dollars	
	Lower Limit	Upper Limit
50	$93,149	$108,851
75	85,788	116,212
90	74,462	127,538
95	63,988	138,012

TABLE 62

CONFIDENCE LIMITS FOR INSTALLED COST OF CYCLONIC SCRUBBERS FOR GTSP STORAGE

Population Size – 5 Sample Size – 3

Installed Cost = $543,730

Conf. Level, %	Installed Cost, Dollars Lower Limit	Upper Limit
50	$491,266	$596,194
75	442,076	645,384
90	366,385	721,075
95	296,394	791,066

TABLE 63

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR CYCLONIC SCRUBBERS FOR GTSP STORAGE VENTS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor					$ 1,950
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance					7,933
Replacement Parts					
Total Replacement Parts					10,513
Utilities					
Electric Power	$.011/kw-hr				60,975
Fuel					
Pond Water (Make-up)	$.25/M gal				15,000
Water (Cooling)					
Chemicals, Specify					
Total Utilities					75,975
Total Direct Cost					96,371
Annualized Capital Charges					54,373
Total Annual Cost					150,744

REFERENCES

(1) Stern, Arthur C., Air Pollution, Volume III, Academic Press, New York & London, 1968, p. 221.

(2) Waggman, W. H., Phosphoric Acid, Phosphates and Phosphatic Fertilizers, 2nd ed., Reinhold Publishing Corporation, New York, 1952.

(3) Shreve, Norris, The Chemical Process Industries, p. 340.

(4) Grant, H. O., "Pollution Control in a Phosphoric Acid Plant", Chemical Engineering Progress, (Vol. 60, No. 1), January, 1964, p. 53-55.

(5) Annon., "Cleanup Pays Off for Fertilizer Plant", Environmental Science and Technology, (Vol. 6, No. 5), May, 1972, p. 401-402.

(6) Slack, A. V., Phosphoric Acid, Vol. 1, Fertilizer Science and Technology Seives, Marcel Dekker, Inc., N.Y. 1968.

(7) Specht, R. C., and R. R. Calaceto, "Gaseous Fluoride Emissions from Stationary Sources", Chemical Engineering Progress, (Vol. 63, No. 5), May 1967, p. 78-84.

(8) Teller, A. J., "Control of Gaseous Fluoride Emissions", Chemical Engineering Progress, (Vol. 63, No. 3), March 1967, p. 75-79.

(9) English, M., "Fluorine Recovery from Phosphate Fertilizer Manufacture", Chemical and Process Engineering, December, 1967, p. 43-47.

(10) State of Florida Air Pollution Regulations 17-204(6)-1.

(11) Gilbert, Nathan, L. A. Hobbs and W. D. Sandberg, "Adsorption of Hydrogen Fluoride in a Limestone Packed Tower", Chemical Engineering Progress, (Vol. 49, No. 3), March, 1953, p. 120-128.

(12) Yatlov, V. S. and E. N. Pineskaya, Zhur. Priklao Khim., 14, No. 1, 11-13, 1941.

(13) Lunde, K. E., "Performance of Equipment for Control of Fluoride Emission," Industrial and Engineering Chemistry, (Vol. 50, No. 3), March, 1958, p. 293-298.

(14) Ceilcote Corporation, private communication.

(15) Wellman-Powergas Co., private communication.

(16) Farmland Corporation, Lakeland, Florida, private communication.

(17) Patty, Frank A., Industrial Hygiene and Toxicology, Volume I, Interscience Publishers, Inc., New York, N.Y., 2nd Ed., 1958, p. 289-293.

2. FEED AND GRAIN INDUSTRY

Slightly more than 86% of the cereal grains raised in the United States today and consumed domestically goes for the feeding of livestock and poultry.(1) The feed industry combines cereal grains and mixtures of grain with other natural products and synthetics to produce what are called "formula" or "mixed" feeds. Millions of farmers still feed some straight grain, sometimes fortified with an oil meal, to their livestock. However, the trend is toward wider use of commercially formulated manufactured rations.

Corn gluten* meal is recognized as the first dry protein cereal by-product to have been fed widely to livestock. Corn gluten meal is a by-product of the wet milling of corn. In this process, whole grain corn is steeped in a dilute sulfurous acid solution and subsequently separated into starch, fiber, gluten, and corn germ**.

Some of the gluten meal is mixed with coarse and fine fiber (hulls and bran***) and other low-protein by-products to make a lower-grade feed than the meal itself. Drying and handling of this final feed are frequently done in a flash drying system.

Other by-product feedstuffs are frequently recovered in the manufacture of the main food product. Whenever the by-product is high in protein value, it is dried, ground, and sold as a high-protein feed, or mixed with low protein by-products (chaff, hulls, bran, etc.) to increase their feed value. Other important high-protein feeds are oil-seed meals (soybean, cottonseed, flaxseed and safflower seed), dried distillers' solubles****, fish meal, and various animal residues (blood, tankage*****, etc.). Table 64.(1) presents the composition of several feedstuffs used in the manufacture of feed.

Feed manufacturing is a comparatively simple operation. Cereal grains, oil seed meals, packinghouse by-products, and occasionally minute quantities of vitamins, molasses, and other ingredients, are combined to produce rations. If every trace element and vitamin additive were individually counted and added to the total number of basic ingredients such as the grains, a formula could easily contain 50 or more components.

---

*Gluten is a term which describes the protein derived from grains, principally corn or wheat.

**Germ is the seed part of the grain.

***Bran is the skin or husk of the grain.

****Dry distillers' solubles are by-products of the brewing and distilling industry.

*****Tankage refers to the residual solids created in the wet rendering of animal carcasses.

TABLE 64

COMPOSITION OF TYPICAL FEEDSTUFFS[1] USED IN MANUFACTURE OF FEED

Description	Crude Fiber	Cell Walls	Acid Deterg. Fiber	Lignin	Crude Protein	Dry Matter Ruminant	Digestibility Non-ruminant
			Dry Matter Basis				
Corn (dent) grain	2.2	10.1	2.7	0.6	9.8	86	
Wheat grain	2.3	12.4	3.1	0.7	11.6	85	
Barley grain	7.0	20	6.3	1.0	12.7	80	
Oats grain	12	31	16.4	2.8	13.6	75	
Soybean meal	6.4	12.0	9.0	0.3	44	84	
Sesame meal	6.9	16.8	9.8	2.0	52	77	
Linseed meal	9.0	25	16.4	0.7	38	81	
Cottonseed meal	12.4	27	18.0	6.6	44	19	
Safflower meal	35	58	41	13.7	24	50	
Cooked feathers	1.5	19.5	12.2	–	96	–	70
Raw feathers	1.3	9.0	65.0	–	96	–	< 10
Tankage	3.7	25	3.6	–	67	–	68
Rice bran	13	24.1	15.9	4.3	15	61	
Malt sprouts	12.5	45	16	1.1	26	77	
Wheat bran	11	47	12.1	4.0	17.4	73	
Dried beet pulp	21	54	33	2.5	10.0	76	
Alfalfa hay (early)	23.5	40	25	5.3	20	62	50
Alfalfa hay (late)	39	55	40	9.0	14	53	
Orchardgrass (early)	24	52	27	2.7	24	72	38
Orchardgrass (late)	35	70	40	4.7	11	57	20
Timothy	32	65	43	7.0	10	50	
Coastal Bermudagrass	32	76	38	5.6	6.4	46	
Soybean hulls	28	63	45	2.0	15	68	
Soybean straw	43	67	53	12.0	11	44	
Barley straw	38	72	45	5.2	2.6	54	
Wheat straw	42	82	53	7.6	3.2	36	
Corn cobs & husk	32	83	45	5.0	2.3	50	
Rice hulls	43	81	66	14	3.0	< 10	

Table 65[1] provides illustrations of rations and supplements for one class of swine, the brood sow. The formula under "ration" is for a complete diet, ready for the sow. If the "supplement" formula is used, a ton of complete feed would be prepared by adding 1,100 lb of ground corn and 400 lb of ground oats to 500 lb of the supplement. The supplement type of feed is usually used in the feeding of beef cattle. It provides sufficient protein to balance that which is in the grain and roughage which the cattle eat. Table 66[1] gives four formulas for beef cattle supplements, three providing 32% protein and one with 65% protein. The major difference between these supplements is that one is based on all natural protein, while the other three include a non-protein source in the form of urea.

The feed formulas presented in Tables 65 and 66 are representative of those being manufactured in the United States today. The composition of feed for other animals and poultry vary from these only in relation to the differences in their requirements for various nutrients.[1]

In both the manufacture of feed and the manufacture of flour, a process called milling is employed. However, feed milling and flour milling are two entirely distinct processes. The function of feed milling is to reduce the cereal grains and oilseed meals to the desired size and consistency of particle, and then to mix this ground material with other ingredients to produce scientifically formulated rations. The reduction in size of the cereal grains and other constituents is usually performed in a grinding device, such as a hammer mill.

Flour and products made from flour have been used by man for centuries. The bland taste of food made from cereal grains has much to do with this fact. The first articles made from flour were unleavened breads. Later, the Egyptians introduced yeast leavened bread.

The primary reason for the continued use of cereal grains as a food is not taste, but the ease of growing cereals and their stability in storage. The power of drying as a means of preserving cereal grains was discovered by early man. He used sun drying techniques to avoid spoilage and antique milling techniques to improve the taste of the food products.

The original millstones, which have been found in ruins that date back to 4000 B.C., were essentially a crude mortar and pestle arrangement designed to crush wheat into smaller particles. Then it was discovered that moving one millstone on another in a semicircular fashion was a more efficient means of reducing grain to flour. Power to run such mills often came from animals. The water wheel was invented around 100 B.C., and wind and water then became

TABLE 65

RATIONS AND SUPPLEMENTS FOR BROOD SOWS(1)

	Ration	Supplement
Ground corn	1,110 lb	–
Ground oats	400	–
Middlings	50	200
17% dehy. alfalfa meal	50	200
50% meat & bone meal	75	300
60% fish meal	25	100
44% soybean meal	250	1,030
Dicalcium phosphate (18%)	10	40
Calcium carbonate	20	80
Salt	8	30
Trace mineral mix	1	4
Hog vitamin premix	5	20
Total batch	2,004 lb	2,004 lb

CALCULATED NUTRIENT CONTENT

	Ration	Supplement
Protein, %	16.2	36.2
Fat, %	3.6	2.6
Fiber, %	5.6	7.3
Calcium, %	1.02	4.0
Vitamin A, IU/lb	2,750	10,000
Phosphorus, %	0.62	1.65
Vitamin D_2, USPU/lb	375	1,500
Riboflavin, Mg/lb	2.4	7.7
Niacin, Mg/lb	23.3	66.5
Pantothenic acid, Mg/lb	8.9	26.0
Choline, Mg/lb	530	1,145
Vitamin B_{12}, Mcg/lb	12.5	51.0

GUARANTEED ANALYSIS

	Ration	Supplement
Crude protein, min %	16.0	36.0
Crude fat, min %	3.5	2.5
Crude fiber, max %	6.0	8.0
Calcium (Ca), max%	Not Needed	4.5
Calcium (Ca), min %	Not Needed	3.5
Phosphorus (P), min %	Not Needed	1.6
Iodine (I), min %	Not Needed	0.0006
Salt (NaCl), max %	Not Needed	2.0
Salt (NaCl), min %	Not Needed	1.0

TABLE 66

BEEF CATTLE FEED SUPPLEMENTS(1)

Ingredients	32% Cattle Supplement All Natural Protein	32% Cattle Supplement with 4% Urea	32% Cattle Supplement with 6% Urea	65% Cattle Supplement with 20% Urea
44% Soybean meal	1,320	600	60	50
Wheat middlings	100	750	450	150
Corn gluten feed	100	–	800	–
17% dehy. alfalfa	150	200	200	600
Urea (45%)	–	80	120	400
Dicalcium phosphate	30	30	40	150
Calcium carbonate	100	110	100	60
Salt	50	50	50	80
Trace mineral mix	5	5	5	10
"Dry" molasses product	–	–	–	400
Vitamin A	2	2	2	4
Vitamin D_2	1	1	1	2
Molasses (wet)	150	180	180	100
Total batch, lb	2,008	2,008	2,008	2,006

CALCULATED NUTRIENT CONTENT

Protein, %	32.2	32.2	32.4	64.4
Equivalent protein, %	–	11.24	16.86	56.20
Fat, %	1.05	2.1	2.0	1.04
Fiber, %	7.0	7.5	8.2	11.3
Calcium, %	2.58	2.72	2.73	3.36
Phosphorus, %	0.80	0.81	0.85	1.51
Vitamin A, IU/lb	13,600	13,600	13,600	27,200
Vitamin D_2, IU/lb	2,000	2,000	2,000	4,000

GUARANTEED ANALYSIS

Crude protein, min %	32.0	32.0	32.0	64.0
This includes no more than -% equivalent crude protein from non-protein nitrogen	–	11.5	17.0	56.5
Crude fat, min %	1.0	2.0	2.0	1.0
Crude fiber, max %	8.0	8.5	9.0	12.5
Calcium (Ca), max %	3.0	3.0	3.0	4.0
Calcium (Ca), min %	2.5	2.5	2.5	3.0
Phosphorus (P), min %	0.8	0.8	0.8	1.5
Iodine (I), min %	0.0006	0.0006	0.0006	0.001
Salt (NaCl), max %	3.0	3.0	3.0	5.0
Salt (NaCl), min %	2.5	2.5	2.5	4.0

the prime sources of energy for milling until comparatively recent times.

The function of flour milling is to separate the starchy endosperm, which is the part that is ultimately made into flour, from the bran and germ of the grain. The bran is composed of several outer coverings of the grain. The germ is the embryo of the new plant.

There are many manufacturing processes used within the Feed and Grain Industry. The scope of this contract limits detailed consideration to the following processes:

1. Feed Milling

2. Flour Milling

3. Feed Flash Drying

FEED MANUFACTURING

Feed manufacturing involves the combining of cereal grains, oilseed meals, and packinghouse and other by-products to produce rations most beneficial to the production of meat, milk, eggs or fur. Molasses, vitamins, drugs and other ingredients are occasionally used in minute quantities to further improve upon the formulation. A flow diagram of a simplified feed manufacturing facility, commonly referred to as a feed mill, is shown in Figure 51.(2)

Feed materials are usually shipped to the feed mill by truck or rail. The medium-size and large-size mills sometimes receive their required grain by barge. Grain is unloaded by gravity, by air or by hand. Frequently, railroad boxcars are unloaded with power shovels and hand labor. Positive or negative pressure air systems are generally used for unloading barges.

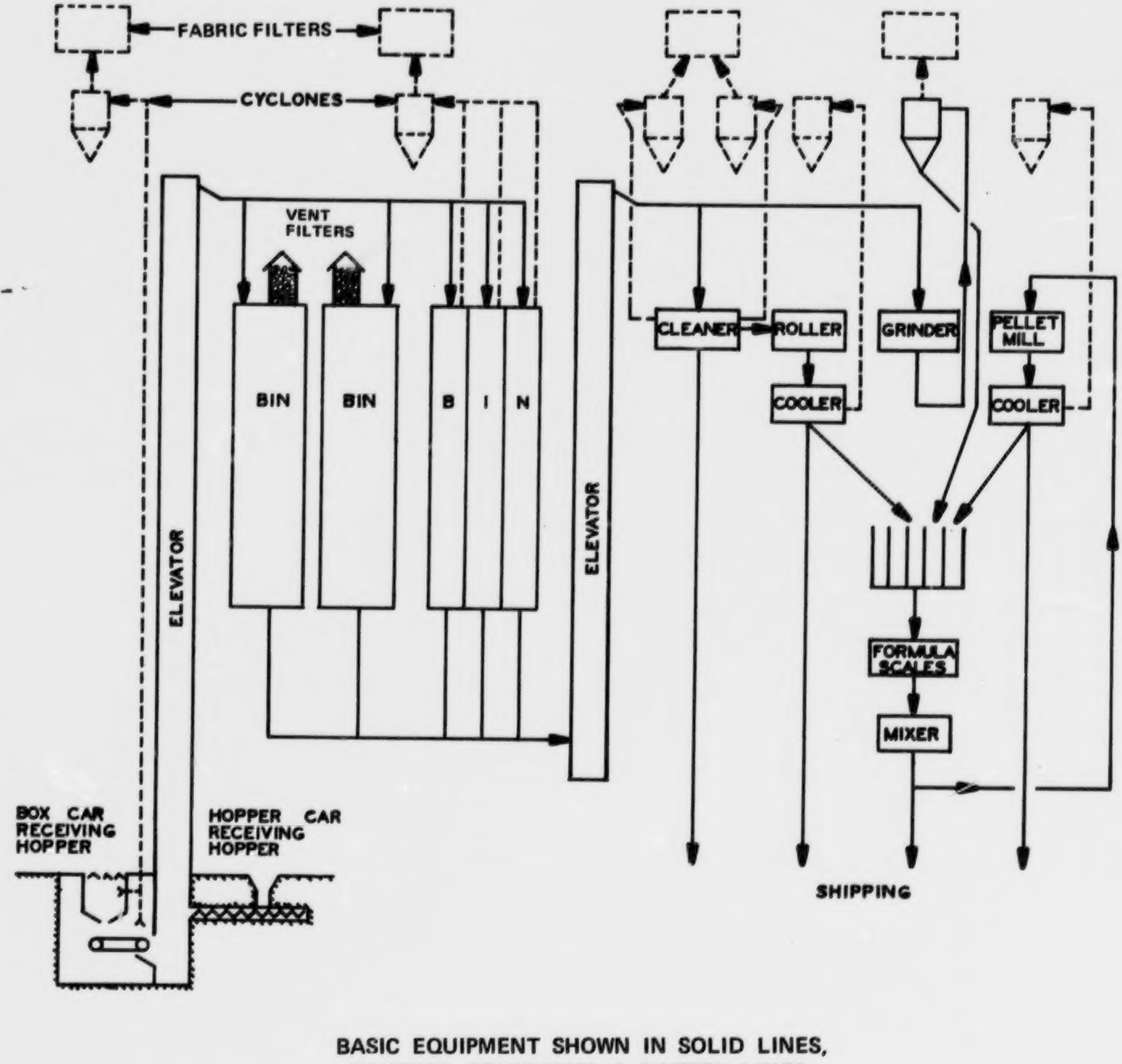
The movement of grain within the mill is usually by bucket elevator and gravity.

Grain is usually stored in concrete bins. Its moisture content usually does not exceed 15%, and 12% moisture content is preferred. If necessary, the grain can be dried in a separate structure adjacent to the mill.

From the storage bins, whole grains are conveyed to cleaning, rolling, grinding, and other plant processes. The processed grains may be shipped to consumers or held for feed formulation. Finished feed formulas can be

FIGURE 51

SIMPLIFIED FLOW DIAGRAM OF A
FEED MILL


FABRIC FILTERS
CYCLONES
VENT FILTERS
BIN
BIN
B
I
N
ELEVATOR
ELEVATOR
CLEANER
ROLLER
GRINDER
PELLET MILL
COOLER
COOLER
FORMULA SCALES
MIXER
SHIPPING
BOX CAR RECEIVING HOPPER
HOPPER CAR RECEIVING HOPPER

BASIC EQUIPMENT SHOWN IN SOLID LINES,
CONTROL EQUIPMENT IS DOTTED LINES.

prepared in the form of finely ground mash*, pellets, or mixed mash and pellets.

Grinding

The basic grinding device is the hammer mill. The mill's grinding chamber houses rows of loosely mounted swinging hammers and plates of hardened steel. These hammers pulverize the grain by striking it sharply. The pulverized grain is ejected from the mill when it is ground finely enough to pass through a screen which is part of the mill. The screens have openings ranging from 1/8 in. upwards to 1/2 in. The fineness of grind is determined by the type of feed in which the ground material is to be used.

Other types of mills used in feed production include the attrition mill and roller mills. In the attrition mill, revolving grinding plates pulverize the grain to produce a finely textured soft product. Roller mills include pairs of corrugated metal rolls, one of which revolves at a speed 2 to 4 times that of the other. The grain is forced between the rolls and sheared by the corrugation.

A fan is required to convey the ground material away from the grinding unit, whether it be a hammer mill, attrition mill or roller mill.

Cleaning

Scalpers**are one of several devices used to remove coarse, non-edible material from the product flow before it reaches the mixer. Separators, which consist of reciprocating sieves which classify grains of different sizes and textures, perform a similar function.

Permanent or electric magnets are usually installed ahead of the grinder for the removal of stray pieces of wire or metal.

Mixing

There are three basic types of feed mixers:

*Mixture of ground grain and water.

**Revolving wire cage containing curved, internal baffles.

1. Vertical

2. Horizontal

3. Drum

The vertical mixer is the most widely used in the smaller plants. It can range in capacity from 500 lb to 7 tons and is usually a batch operation.

The ground grain and other ingredients, which may include a protein source, vitamins, minerals, and drugs, are measured into the mixing cylinder. The screws or paddles inside the cylinder are then operated for the necessary time to establish the proper blending of the mixture.

A dry mash feed is ready for packaging, bulk delivery, or storage after it has been mixed.

Pelleting

A major advantage of pellets is that all the components are bonded together so that the bird or animal consumes the proper balance of nutrients intended for it.

The manufacture of pellets begins with the blending of molasses or fat with the dry mash. A binding agent is also incorporated into the mixture. Live steam is added to the dry mixed feed as it enters the pellet mill. After the pellets are formed and compressed, they are released into a cooling system which hardens the pellets. Broken pellets and milling fines are returned to the pellet mill for reprocessing.

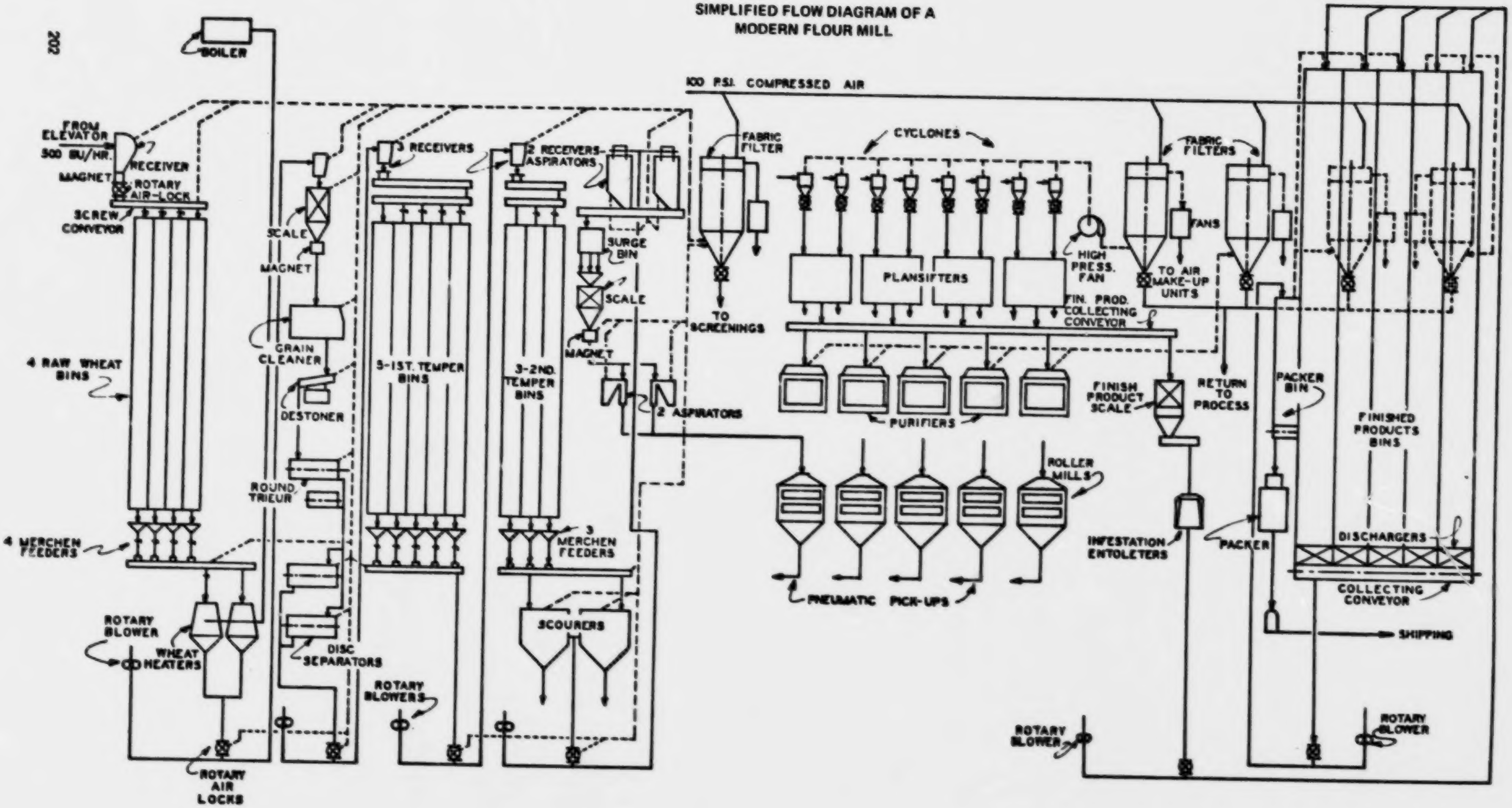
FLOUR MILLING

Not all cereal grains are pulverized to a flour. Rice and barley are milled in such a way that only the husk is removed. Oats are rolled into a flattened form. The following is a description of the milling process used for the conversion of wheat into flour.

A simplified flow diagram of a modern flour mill is shown in Figure 52.(3) Wheat is stored at the mill in large bins. Most mills store at least 6 weeks supply, with some having over 42 weeks storage capacity. On arrival, wheat contains foreign impurities such as sticks, stones, string, rodent contamination,

FIGURE 52

SIMPLIFIED FLOW DIAGRAM OF A MODERN FLOUR MILL


FROM ELEVATOR
300 BU/HR.
RECEIVER
MAGNET
ROTARY AIR-LOCK
SCREW CONVEYOR
4 RAW WHEAT BINS
4 MERCHEN FEEDERS
ROTARY BLOWER
WHEAT HEATERS
ROTARY AIR LOCKS
BOILER
SCALE
MAGNET
GRAIN CLEANER
DESTONER
ROUND TRIEUR
DISC SEPARATORS
3 RECEIVERS
5-1ST. TEMPER BINS
ROTARY BLOWERS
2 RECEIVERS-ASPIRATORS
3-2ND. TEMPER BINS
3 MERCHEN FEEDERS
SCOURERS
SURGE BIN
SCALE
MAGNET
2 ASPIRATORS
100 PSI. COMPRESSED AIR
FABRIC FILTER
TO SCREENINGS
CYCLONES
PLANSIFTERS
HIGH PRESS. FAN
FIN. PROD. COLLECTING CONVEYOR
PURIFIERS
ROLLER MILLS
PNEUMATIC PICK-UPS
FABRIC FILTERS
FANS
TO AIR MAKE-UP UNITS
FINISH PRODUCT SCALE
RETURN TO PROCESS
INFESTATION ENTOLETERS
ROTARY BLOWER
PACKER BIN
PACKER
FINISHED PRODUCTS BINS
DISCHARGERS
COLLECTING CONVEYOR
SHIPPING
ROTARY BLOWER

NOTE: DOTTED EXHAUST LINES CONNECT ALL ITEMS OF EQUIPMENT TO FABRIC FILTERS etc. which are removed before wheat goes into storage. Dust and smaller pieces of foreign matter are removed in the cleaning done just prior to milling.

If the wheat is too wet, it is dried before going into storage. If the moisture of the grain is above 14.5%, heat production results from the growth of microorganisms in the grain. Hot air is used for the drying of wheat. Evaporated moisture is carried away by air currents.

Cleaning Prior to Milling

As grain is needed for milling, it is withdrawn from storage and sent through a cleaning operation.

Prior to entering the cleaning house from storage, the wheat is weighed. Then it goes through the same type of separator used in cleaning prior to storage, except the separator is set to remove fine impurities and dust. Light impurities such as wheat chaff are removed by air currents. Next, the wheat passes through magnetic separators which remove bits of metal which inadvertently found their way into the wheat mix.

Then additional aspiration is provided, from which the wheat enters a machine designed to remove cereal grains other than wheat and to remove foreign seeds, particularly weed seeds. The unwanted seeds are caught in specially designed pockets in a revolving metal plate, and are lifted up and carried away, while the wheat passes through the machine.

Another cleaning operation is performed in a scourer. Scouring, a dry operation, involves the removal of dirt on the surface of the wheat by friction. Sometimes a light washing is performed prior to scouring in order to toughen the bran coat.

The final cleaning step is a water wash, which dissolves the dirt and permits stones and bits of metal to sink. The washer adds about 1% to the moisture content of the wheat. In a typical washer, wheat is conveyed through a trough of water where dirt is floated away and stones sink. After washing, excess surface moisture is removed from the wheat by centrifugal forces.

Tempering and Conditioning

Tempering refers to the addition of water to the bran and endosperm. The bran becomes tough and rubbery while the endosperm becomes softer, which improves the milling efficiency. Conditioning employs a controlled heating process in addition to the moistening procedure.

Wheats can be broadly classified into the classes of hard wheat and soft wheat according to their milling characteristics. Hard wheats require about 20% more horsepower to grind to a given size. When hard and soft wheats are found in the same wheat mix, the problem evolves into one of overgrinding the soft wheat or undergrinding the hard wheat. If hard wheats are maintained in the presence of moisture under the proper conditions and for appropriate periods of time, they grind more like soft wheats.

Tempering, as it is still practiced in the United States, involves adding water to grain to raise the moisture to 15 to 19% for hard wheats and 14.5 to 17% for soft wheats. An excessive amount of water is added to the grain, which lies in tempering bins. The unabsorbed surface moisture is drained from the tempering bins, and the wheat is allowed to lie in the bins for periods of 18 to 72 hours. Usually, tempering is done in successive steps since it is impractical to add more than a few percent of water to wheat at one time.

Conditioning, in contrast to tempering, always involves the use of heat. Quick diffusion of water into the endosperm as well as the bran is the purpose. Normally, a temperature of 115°F cannot be exceeded without causing detectable changes in the baking quality of the flour.

Wheat conditioners consist of four sections. The first section heats the wheat to the proper temperature. The second adds moisture and holds the wheat for the proper time. The third section cools the wheat to room temperature. The final section allows time for the moisture in the wheat to equilibrate prior to milling.

The wheat goes through the first three sections in 1.5 hours or less. The wheat is held in the fourth section for 8 to 18 hours, with the longer times being used for the harder wheats.

Grinding

The purpose of flour milling is first to separate the endosperm from bran and germ in chunks as large as possible and then to reduce the size of the endosperm chunks to flour-sized particles through a series of grinding steps.

The grinding of wheat is done between pairs of rolls. These rolls move in opposite directions and at different rates of speed. They are set with an appreciable gap between them, such that they do not grind the wheat primarily by crushing, but rather by shearing forces which are relatively gentle. The rapidly moving roll runs about 2.5 times faster than the slower one and at speeds from 250 to 450 rpm.

The roller milling area is divided into the break section and the reduction section. The bran is broken open and the endosperm is milled away in the break section. This section usually involves four or more sets of rolls, each taking stock from the preceding one. After each break, the mixture of free bran, free endosperm, free germ, and bran containing adhering endosperm is sifted. The bran having endosperm still attached goes to the next break roll, and the process is repeated until as much endosperm has been separated from the bran as is possible.

The surface of break rolls is always fluted to obtain the necessary grinding effect. The number of saw-tooth flutes increase from 10 to 12 flutes per inch on the first break, to about 28 flutes per inch on the fourth break.

A sifting system (called scalping) follows each set of break rolls. The scalping system is a combination of a sieving operation (plansifters) and air aspiration (purifiers).

A plansifter consists of flat sieves piled in tiers, one above the other. The action of the sifter is rotary in a plane parallel with the floor. As the sifter moves in a 3.5 in. circle, the small-sized particles travel across the sieve to a collecting trough and are removed. Large pieces of bran are first removed and are sent to the next break roll. Bran and germ are scalped off the finer sieves. Below these are the flour sieves. The resulting flour and endosperm chunks (middlings) still contain minute size bran particles. These are removed by sending the product through a purifier where air currents carry the bran away.

A purifier is a long oscillating sieve, inclined downwards, with air currents passing through the sieve in the direction from floor to ceiling. The buoyancy effect of the air causes the flour to stratify into endosperm chunks of different sizes. Thus, a purifier is used to make a coarse separation of middlings as well as for the removal of bran. The endosperm chunks fall through the sieve openings and are sent to the appropriate reduction rolls. The overs are a composite of bran and bran plus endosperm. These go back to the break rolls or to millfeed stock. Aspirated materials go to millfeed.

For a flour roll break section, a scalping system containing 12 purifiers is normal. The fine middlings from the first, second, third and fourth breaks normally go through a double purification step while only one purification treatment is needed for the coarser middlings.

The reduction system consists of two parts, roll mills and sifting machines. The major difference from the break system is that the surface of the reduction mills is smooth rather than grooved. The reduction rolls reduce endosperm middlings to flour size and facilitate the removal of the last remaining particles of bran and germ.

In a reduction system the roll stands are divided into coarse rolls and fine rolls. Coarse and fine is in reference to the setting of the rolls, whether they are set wide to produce coarse grinding or close to produce fine grinding. The middlings produced from the coarse rolls are sent to the fine rolls to be ground to flour.

After each reduction, the resulting product is classified by particle size in a sifter. Oversized material is sent back to the reduction rolls for further processing. Normally there is at least one sifting device for every reduction roll stand.

Conveying System

Older mills depend upon gravity for the transport of flour stock from machine to machine. The wheat and flour is moved to the top of the mill by bucket elevators, and then the material flows by spouts to the rolls and to the sifters.

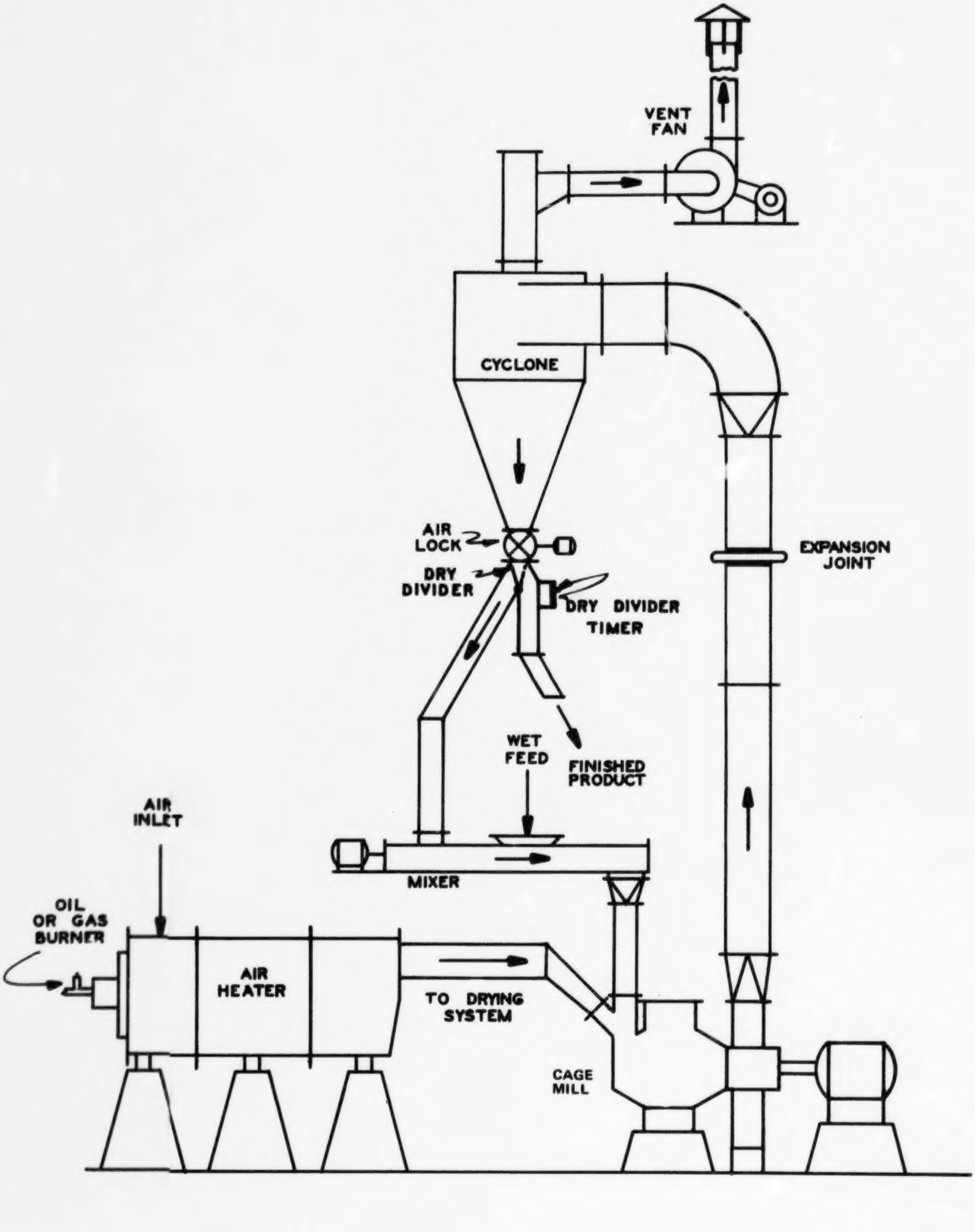
Bucket elevators are dusty and they provide a place for insects to grow. Consequently, flour mills are abandoning bucket elevators and gravity spouts, and are converting to air conveying. Powerful suction fans provide the negative pressure. The air intake is through the roller mills. This air movement has the desirable effect of keeping the rolls and the flour cool during grinding. The simplified flow diagram of a modern flour mill shown in Figure 2 includes an air conveying system.

FEED FLASH DRYING

Flash drying systems, as applied to the feed industry, are usually designed to accomplish drying with disintegration. Such a system is shown in Figure 53.(2) This is in contrast to simultaneous drying and grinding operations. Since the desired particle size has been established in the original wet feed, the dried material is simply disintegrated, i.e., separation of loosely packed materials.

FIGURE 53

FLASH DRYING COMBINED WITH DISINTEGRATION


VENT FAN
CYCLONE
AIR LOCK
DRY DIVIDER
DRY DIVIDER TIMER
EXPANSION JOINT
WET FEED
FINISHED PRODUCT
AIR INLET
MIXER
OIL OR GAS BURNER
AIR HEATER
TO DRYING SYSTEM
CAGE MILL

Disintegration is usually accomplished with a cage mill.

In a flash dryer, or pneumatic conveyer dryer, moisture is removed by dispersing the material to be dried in a hot gas zone followed by conveying at high velocities. The dispersion is usually accomplished with a cage mill, a device designed to break up loosely packed materials, or an imp mill for more tightly compacted materials. The imp mill is a high-speed impact pulverizer.

The dryer system consists of a furnace or other source of hot gases, a dispersion device, a duct through which the gases convey the material, and a collection system for removing the dry product from the gas stream.

Frequently, the wet feed must be conditioned by mixing with recycled dry product in a mixer. This is required in order to achieve good dispersion in the cage mill. Filter cakes, the form of many by-products of wet milling processes, have very unfavorable drying characteristics. They are usually conditioned by mixing with previously dried material.

NATURE OF THE GASEOUS DISCHARGE

There are many sources of air pollution in the feed and grain industry. Table 67 (4) presents the more common sources.

The air pollution problems with regard to receiving, handling, and storing operations are generally due to the fine dusts found in field run grains. Fine particles are produced from the grain itself, by abrasion in handling and storing. The fine dust may include the soil in which the grain was grown, or fine particles originating from weeds or insects. Therefore, because of these possible sources of fine dust, no reliable prediction of the kind and amount of dust in any specific shipment of field run grain can be made.

Table 68 (2) presents the particle size distribution of dusts generated during the unloading of barley from a boxcar. The barley was received in a deep hopper equipped with a control hood. A blower carried the dust picked up by the control hood to a cyclone where the larger particles dropped out and were collected in a sack (sample no. 1). The cyclone was vented to a fabric filter, which collected the finer material in a hopper (sample no. 2).

Grain grinding, which is commonly done in feed mills with a hammer mill, generates a large amount of fine dust. A fan conveys away the ground materials from the hammer mill and discharges it into a cyclone. Typical dust losses from a hammer mill cyclone, along with the dust losses from some grain cleaner

TABLE 67

SOURCES OF AIR POLLUTION IN FEED AND GRAIN INDUSTRY[4]

Source	Emission Type
Grain Unloading	Fine Dust
Grain Handling & Transport	Fine Dust
Grain Cleaning	Fine Dust
Grain Storage	Fine Dust
Grain Grinding	Fine Dust
Pelletizing of Feed Mixtures	Fine Dust (sometimes Odor)
Feed Loading into Trucks & Rail Cars	Fine Dust
Flour Milling	Very Fine Dust
Loading of Middlings & Chaff from Flour Milling into Rail Cars	Fine Dust
Loading Grain Dust Cars	Fine Dust
Oil-Seed Expelling	Odor
Flash Drying	Dust, Vapors & Odor
Barley Malt Toasting	Odor

TABLE 68[2]

PARTICLE SIZE ANALYSIS OF THE PRIMARY CYCLONE CATCH AND THE SECONDARY FABRIC FILTER CATCH OF DUST FROM A RAILROAD RECEIVING HOPPER HOOD CONTROLLING THE UNLOADING OF A BOXCAR OF FEED-TYPE BARLEY[a]

	Particle Size Distribution by Weight	
Particle Size, μ	Sample No. 1 Cyclone Bottoms, %	Sample No. 2 Fabric Filter Hopper, %
0 to 5	0.9	4
5 to 10	0.9	25
10 to 20	3.9	66
20 to 44	9.3	5
44 to 74	12.9	0
74 to 149	16.2	0
149 to 250	5.4	0
Over 250 (60 mesh)	50.5	0

(a)Specific gravity of both samples was 1.8.

cyclones, are presented in Table 69.[2]

During the drying of a high-protein feed, the air contaminants that may be emitted are dusts, vapors, and odors. If the material is agitated or stirred during the drying process, such as in a flash drying system, dust emissions are likely to be a problem. The nature of the emissions is determined by the material being dried and the operating conditions of the drying system.

A typical flash drying system will reduce the moisture content of wet feed from 60% down to 30%. For a 20,000 lb/hr feed rate, this is equivalent to removing 8,570 lb/hr of moisture as a vapor. The gas flow associated with this rate of water removal is 20,000 ACFM at 220° F.

In the modern flour mill the product flows by gravity from the top of the building, through the various machines, to the ground floor. At the completion of each phase of processing, the material is relifted to the top floor by pneumatic conveyors for further processing. Many passes up and down the building are required to produce the finished product. A considerable amount of piping and auxiliary equipment is required in connection with these pneumatic conveying systems. The conveying systems are typically divided up according to their function as follows:

1. System for delivery of raw wheat from a nearby grain elevator
2. Pneumatic relift conveyors for the wheat cleaning and tempering processes
3. Pneumatic conveyor for collecting the screenings from the cleaning process
4. Pneumatic relifts for the wheat milling process
5. Pneumatic conveyors for the storage of finished flour and feed

All of the above pneumatic conveying systems are shown in Figure 52. The design of each of these systems is a function of the nature of the material being moved as well as the distance and rate of flow.

At the end of each pneumatic conveying line, the material is fed to a receiver or cyclone. From the receiver or cyclone, the material falls by gravity through rotary air-locks directly into processing machinery.

Each one of the pneumatic conveying systems is equipped with a dust

TABLE 69

DUST LOSSES FROM CYCLONES[2]

Grain	Malted Barley		Feed Barley	Milo*	
Basic Equipment	Grain Cleaner		Hammer Mill	Grain Cleaner	
	Test No. 1	Test No. 2			
Process wt, lb/hr	53,000	50,000	10,350	11,250	
				First Cyclone	Second Cyclone
Exhaust Air Volume, SCFM	2,970	2,970	3,790	3,680	2,610
Dust Concentration, gr/SCF	0.194	0.160	0.488	0.058	0.006
Dust Loss, lb/hr	4.95	4.07	15.8	1.83	0.13

*A sorghum similar to millet and grown for forage collection system. Each of the five dust collection systems uses fabric filters. In the cleaning section, for example, the dust collector aspirates all receivers on the intermediate relift conveyors. In addition it aspirates the screenings suction conveyor, all cleaning machinery, all feeders and all of the screw conveyors in the cleaning section. The other four filters serve other areas of the plant in a similar manner.

Therefore, the gaseous discharges from a modern flour mill should be, except during periods of upset operation, extremely clean. In fact, since the design of most pneumatic conveying systems allows for recycle of the air used in them, there should be no gaseous discharges, except during periods of upset operation.

POLLUTION CONTROL CONSIDERATIONS

Gases discharged from the following pieces of equipment and/or systems will be considered:

1. Feed Grinder

2. Pneumatic Relift System for the flour milling process

3. Flash Drying System

In the cases of feed grinders and pneumatic relift systems, the emission problem is limited to fine dusts. These fine dusts are the result of losses from product recovery cyclones. Since it is desirable to recover the material lost from these cyclones in a dry state the only practical means of control is the fabric filter.

Since the exhaust fan in a flash drying system is usually placed at the product discharge end of the system, the entire system is under negative pressure. This precludes any emissions, except from the final collector where the product is separated from the air stream.

Since the emissions from a flash drying system include odors as well as fine dusts, the pollution control systems of interest are combustion and absorption. The odor levels in the flash dryer specifications are expressed in "odor units". One odor unit per cubic foot is that concentration of odor which is numerically equivalent to its odor threshold. A level of 1500 odor units per cubic foot would require 1500 dilutions with clear, odor-free air to make it just detectable.

SPECIFICATIONS AND COSTS

Specifications for air pollution control systems were written for three sources in the feed and grain industry:

Flour mill pneumatic conveying systems
Feed mill grinders
Feed flash dryers

A fabric filter system was specified for the flour mill pneumatic conveyors. The specifications are shown in Tables 70 and 71. The system was not specified for total control of the mill. It was only designed to handle the air from the cyclonic dust collectors on the pneumatic conveying system which lifts milled grain from the rolls up to the classifying equipment. Other mill emission sources, such as grain cleaning, are controlled by separate fabric filter systems. Cost data are presented on Tables 72 and 73 and on Figures 54 and 55. Confidence limits for the capital cost data are presented on Table 74 and Figure 56.

A fabric filter system was also specified for the feed mill grinder. In keeping with conventional industry practice, the system was designed to handle only the grinder exhaust rather than the total mill exhaust. Specifications are presented on Tables 75 and 76. Cost data are presented on Tables 77 and 78 and on Figures 57 and 58. Confidence limits for the capital cost data are shown on Table 79 and Figure 59.

The emissions from the feed flash dryer consist of both particulate matter and odor. Specifications were written to deal primarily with the odor problem but included a request that suppliers quote the particulate removal efficiency which would be achieved in their system. Both scrubber systems, using buffered potassium permanganate, and thermal combustion systems were quoted.

Specifications for wet scrubber systems are shown on Tables 80 and 81. Cost data are presented on Tables 82 and 83 and on Figures 60 and 61. Estimated removal efficiencies for particulate matter are also shown on Table 82. The primary component of the operating costs for the wet scrubber systems is the consumption of potassium permanganate and borax. The cost of these chemicals represents about 99% of the direct operating cost. These chemical consumption numbers are based upon theoretical reaction calculations rather than upon actual operating data.

Specifications for thermal combustion systems are shown on Tables 85 and 86. Cost data are presented on Tables 87 and 88 and on Figures 63 and 64.

Preceding page blank

TABLE 70

FABRIC FILTER PROCESS DESCRIPTION FOR FLOUR MILLING SPECIFICATION

A fabric filter is to treat the effluent from the product recovery cyclones in a flour mill. These cyclones are an integral part of the pneumatic conveying systems which move the product from the mills to the associated plansifters at higher elevation for classification. Conveying air and solid are separated by cyclone. The air from the top of the cyclones will be routed to the subject fabric filter. The filtered air is recycled back to the pneumatic conveying systems.

The required ductwork from the cyclone collector outlets to the fabric filter will be provided by others. The vendor is to furnish the fabric filter proper, including hoppers equipped with rotary air locks, the required booster fan, ductwork between the fabric filter and fan, and controls.

The entire collection system will be located inside a building, which will be adjacent to a public highway. There is little likelihood of interference caused by other equipment with the location of pollution control equipment.

The fan shall be capable of overcoming the system pressure drop at the design flow rate while operating at no more than 90% of the maximum recommended speed. The system pressure drop should include enough suction for the interconnecting ductwork between the cyclone collector outlets and the control system inlet. Motors shall be capable of driving fans at maximum recommended speed and the corresponding pressure differential at 20% over design flow rate.

TABLE 71

FABRIC FILTER OPERATING CONDITIONS FOR FLOUR MILLING SPECIFICATION

Two sizes of fabric filters are specified at one efficiency level. Vendors' quotations should consist of one quotation for each of the two sizes, with a representation of the efficiency expected for the unit quoted. The efficiency quoted may be better than the efficiency called for in this specification.

	Small	*Large*
Plant Capacity, cwt./day	3,000	10,000
Process Weight, lb/hr	12,500	41,700
Inlet to Fabric Filter		
Flow, ACFM	6,000	20,000
Temperature, °F	70	70
% Saturation	30	30
Dew Point, °F	38	38
Loading, lb/hr	40	134
Loading, gr/ACF	0.778	0.778
Loading, gr/DSCF	0.779	0.779

CASE 1 – HIGH EFFICIENCY

Outlet Solids Rate, lb/hr	0.515	1.72
Outlet Solids Loading, gr/ACF	0.01	0.01
Outlet Solids Loading, gr/DSCF	.01	.01
Collection Efficiency, wt.%	98.7	98.7
Air-To-Cloth Ratio	6/1	6/1

TABLE 72

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR FABRIC FILTERS FOR FLOUR MILLING

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			6,000	20,000
°F			70	70
SCFM			6,000	20,000
Moisture Content, Vol. %			.75	.75
Effluent Contaminant Loading				
gr/ACF			0.778	0.778
lb/hr			40	134
Cleaned Gas Flow				
ACFM			6,000	20,000
°F			70	70
SCFM			6,000	20,000
Moisture Content, Vol. %			.75	.75
Cleaned Gas Contaminant Loading				
gr/ACF			0.01	0.01
lb/hr			0.515	1.72
Cleaning Efficiency, %			98.7	98.7
(1) Gas Cleaning Device Cost			5,537	11,764
(2) Auxiliaries Cost				
(a) Fan(s)			1,409	3,197
(b) Pump(s)			-	-
(c) Damper(s)			233	317
(d) Conditioning, Equipment			-	-
(e) Dust Disposal Equipment			654	1,421
(3) Installation Cost				
(a) Engineering			509	1,018
(b) Foundations & Support			2,197	4,583
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost			10,539	22,300

FIGURE 54

CAPITAL COSTS FOR FABRIC FILTERS FOR FLOUR MILLING

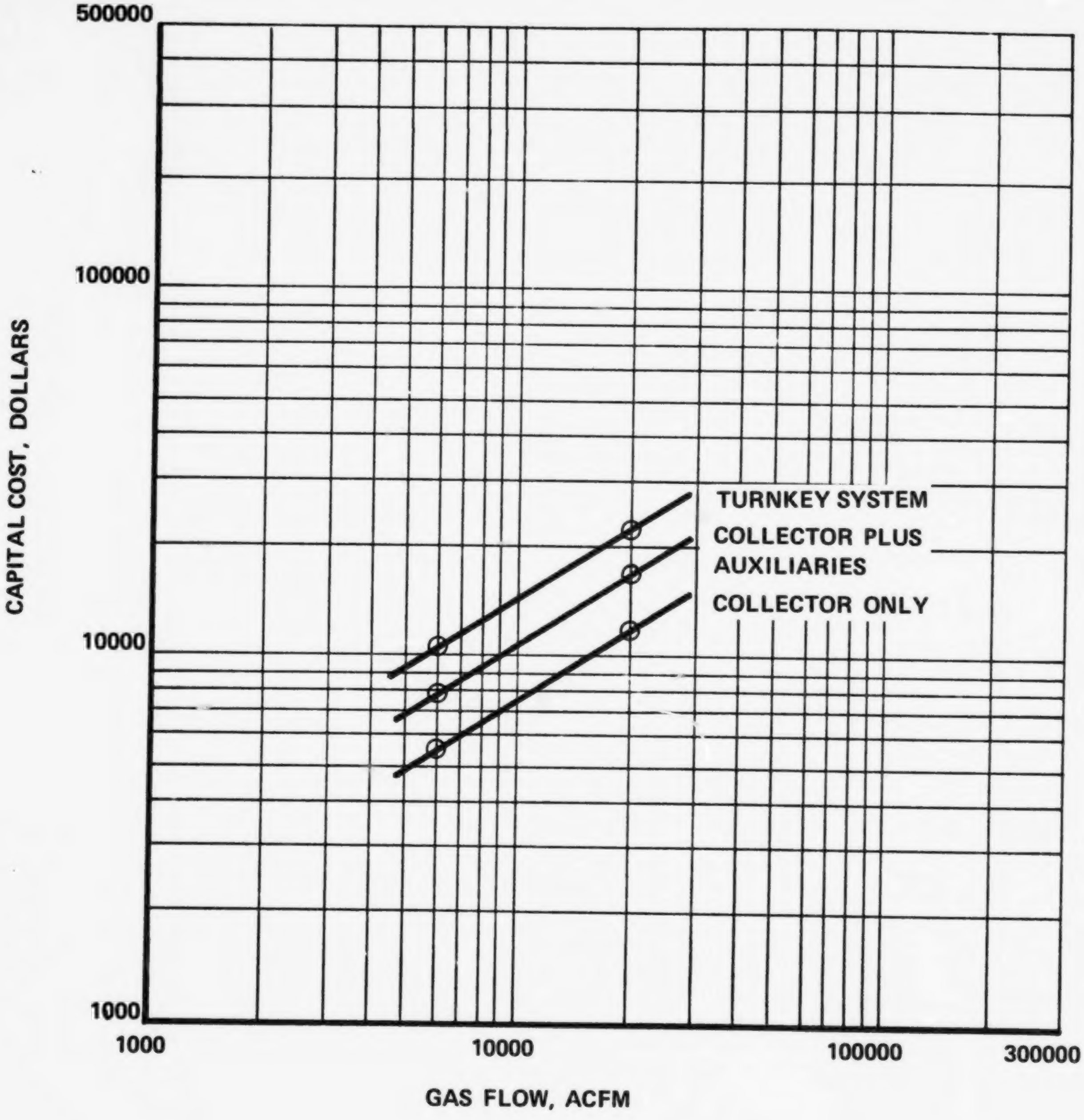


TABLE 73

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR FABRIC FILTERS FOR FLOUR MILLING

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,000				
Operating Labor (if any)					
Operator	$6/hr			-	-
Supervisor	$8/hr			165	165
Total Operating Labor					
Maintenance					
Labor					
Materials					
Total Maintenance				236	554
Replacement Parts				-	-
Total Replacement Parts				350	974
Utilities					
Electric Power	$.011/kw-hr			1,246	3,576
Fuel					
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				1,246	3,576
Total Direct Cost				1,997	5,269
Annualized Capital Charges				1,054	2,230
Total Annual Cost				3,051	7,499

FIGURE 55

ANNUAL COSTS FOR FABRIC FILTERS FOR FLOUR MILLING

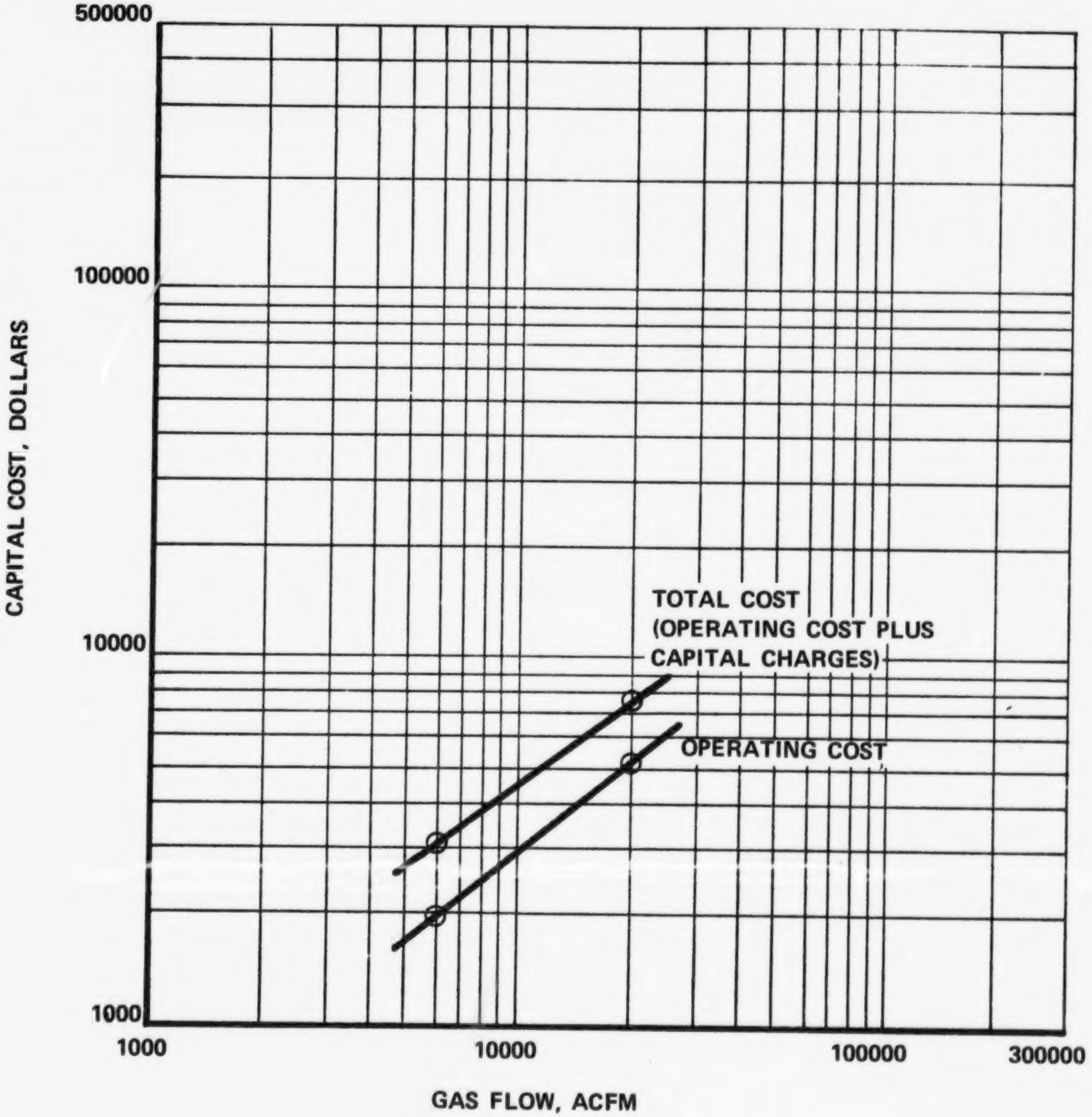


TABLE 74

CONFIDENCE LIMITS FOR CAPITAL COST OF FABRIC FILTERS FOR FLOUR MILLING

Population Size – 20   Sample Size – 3

Capital Cost = $10,540

Conf. Level, %	Capital Cost, Dollars Lower Limit	 Upper Limit
50	$9,921	$11,158
75	9,341	11,738
90	8,449	12,630
95	7,624	13,455

Capital Cost = $22,300

Conf. Level, %	Capital Cost, Dollars Lower Limit	 Upper Limit
50	$21,809	$22,790
75	21,349	23,250
90	20,641	23,958
95	19,987	24,612

FIGURE 56

CONFIDENCE LIMITS FOR CAPITAL COST OF FABRIC FILTERS FOR FLOUR MILLING

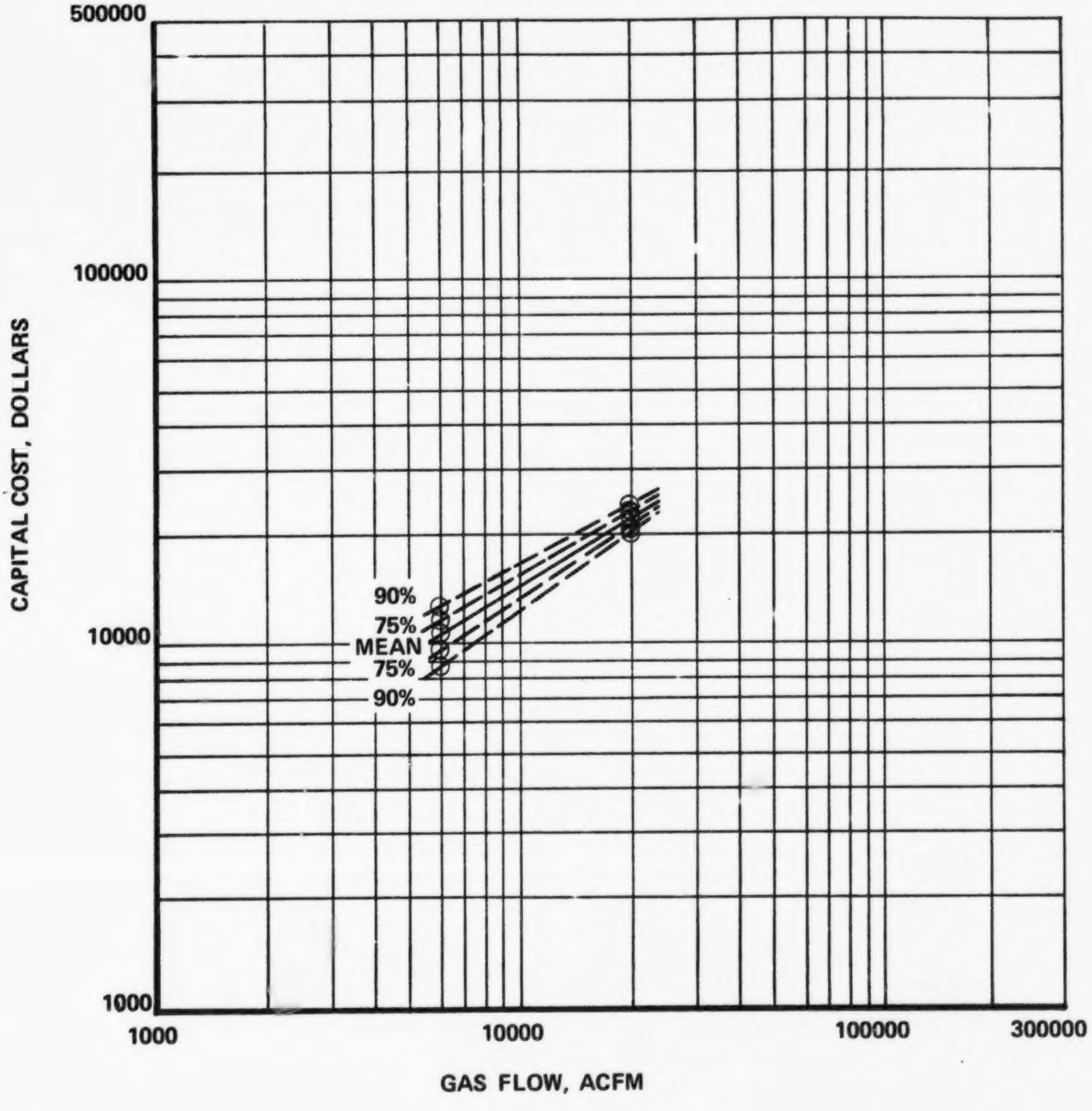


TABLE 75

FABRIC FILTER PROCESS DESCRIPTION FOR FEED GRINDING SPECIFICATION

A fabric filter is to treat the effluent from the product recovery cyclone collectors in a feed mill. These cyclone collectors are an integral part of the feed grinders.

The required ductwork from the cyclone collector outlets to the fabric filter will be provided by others. The vendor is to furnish the fabric filter proper, including hoppers equipped with rotary air locks, the required booster fan, ductwork between the fabric filter and fan, and controls.

The entire collector system will be located inside a building, which will be adjacent to a public highway. There is little likelihood of interference caused by other equipment with the location of pollution control equipment.

The fan shall be capable of overcoming the system pressure drop at the design flow rate while operating at no more than 90% of the maximum recommended speed. The system pressure drop shall be defined to include enough suction for the interconnecting ductwork between the cyclone collector outlets and the control system inlet, i.e., ductwork provided by others than the vendor. Motors shall be capable of driving fans at maximum recommended speed and the corresponding pressure differential at 20% over design flow rate.

TABLE 76

FABRIC FILTER OPERATING CONDITIONS FOR FEED GRINDING SPECIFICATION

Two sizes of fabric filters are specified at one efficiency level. Vendors' quotations should consist of one quotation for each of the two sizes, with a representation of the efficiency expected for the unit quoted. The efficiency quoted may be better than the efficiency called for in this specification.

	Small	*Large*
Plant Capacity, ton/hr	*6*	*24*
Process Weight, lb/hr	*13,200*	*52,800*
Inlet to Fabric Filter		
Flow, ACFM	*4,700*	*19,000*
Temperature, °F	*100*	*100*
% Saturation	*30*	*30*
Dew Point, °F	*64*	*64*
Loading, lb/hr	*18.3*	*73.2*
Loading, gr/ACF	*0.454*	*0.454*
Loading, gr/DSCF	*0.490*	*0.490*

CASE 1 – HIGH EFFICIENCY

	Small	*Large*
Outlet Solids Rate, lb/hr	*0.40*	*1.63*
Outlet Solids Loading, gr/ACF	*0.01*	*0.01*
Outlet Solids Loading, gr/DSCF	*0.011*	*0.011*
Collection Efficiency, wt.%	*97.8*	*97.8*
Air-to-cloth ratio	*6/1*	*6/1*

TABLE 77

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR FABRIC FILTERS FOR FEED GRINDING

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			4,700	19,000
°F			100	100
SCFM			4,420	17,800
Moisture Content, Vol. %			.75	.75
Effluent Contaminant Loading				
gr/ACF			.454	.454
lb/hr			18.3	73.2
Cleaned Gas Flow				
ACFM			4,700	19,000
°F			100	100
SCFM			4,420	17,800
Moisture Content, Vol. %			.75	.75
Cleaned Gas Contaminant Loading				
gr/ACF			.01	.01
lb/hr			.40	1.63
Cleaning Efficiency, %			97.8	97.8
(1) Gas Cleaning Device Cost (incl.Dust Disposal Equip)			5,201	14,395
(2) Auxiliaries Cost				
(a) Fan(s)			1,048	2,476
(b) Pump(s)			-	-
(c) Damper(s)			71	117
(d) Conditioning, Equipment			-	-
(3) Installation Cost				
(a) Engineering (b) Foundations & Support Ductwork Stack Electrical Piping Insulation Painting Supervision Startup Performance Test Other			5,961	15,074
(4) Total Cost			12,281	32,062

FIGURE 57

CAPITAL COSTS FOR FABRIC FILTERS FOR FEED GRINDING

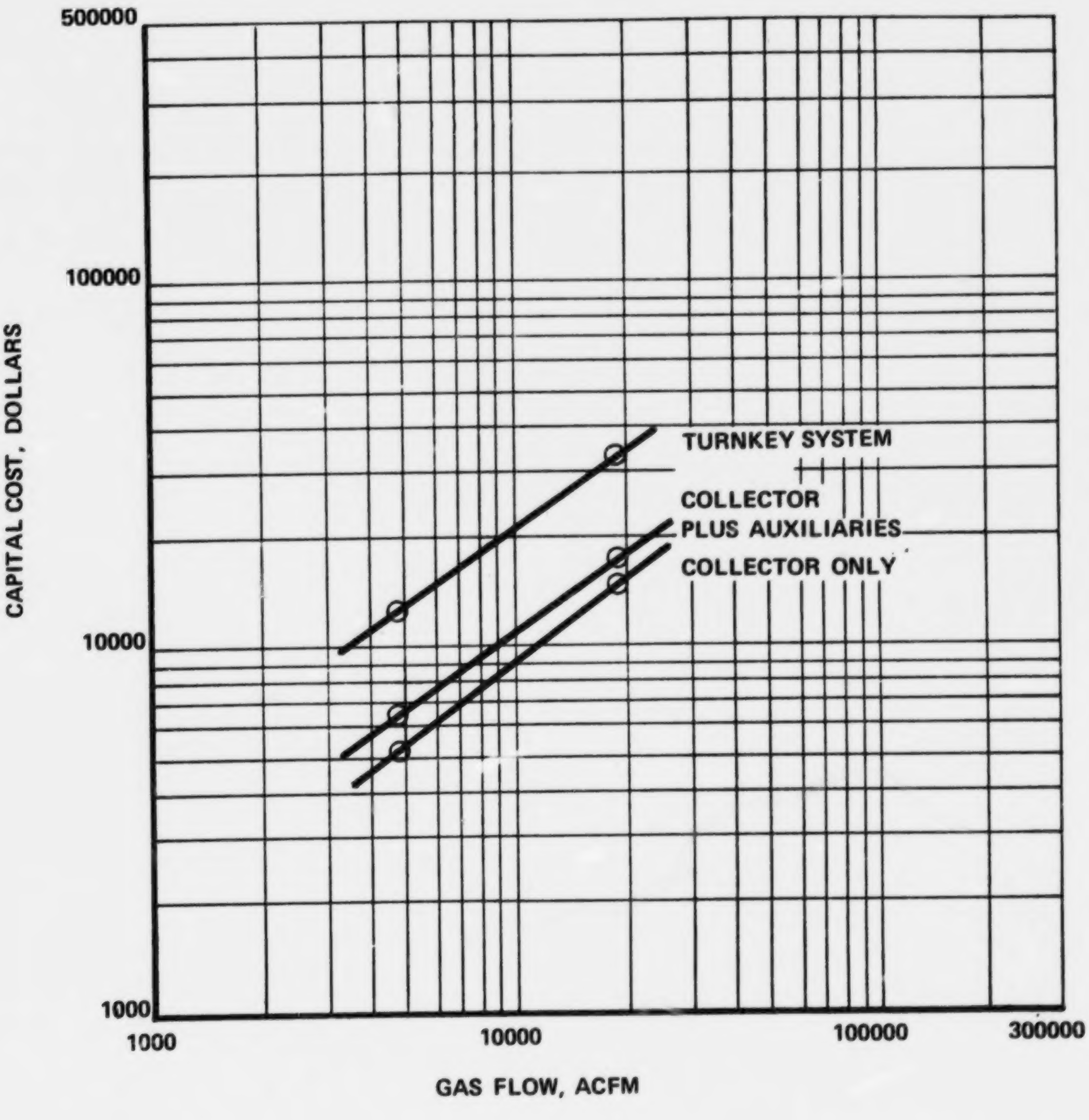


TABLE 78

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR FABRIC FILTERS FOR FEED GRINDING

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,000				
Operating Labor (if any)					
Operator				-	-
Supervisor	$8/hr				
Total Operating Labor				845	845
Maintenance					
Labor					
Materials					
Total Maintenance				134	327
Replacement Parts					
Total Replacement Parts				120	481
Utilities					
Electric Power	$.011/kw-hr			680	2,960
Fuel				-	-
Water (Process)				-	-
Water (Cooling)				-	-
Chemicals, Specify				-	-
Total Utilities				680	2,960
Total Direct Cost				1,779	4,613
Annualized Capital Charges				1,228	3,206
Total Annual Cost				3,007	7,819

FIGURE 58

ANNUAL COSTS FOR FABRIC FILTERS FOR FEED GRINDING

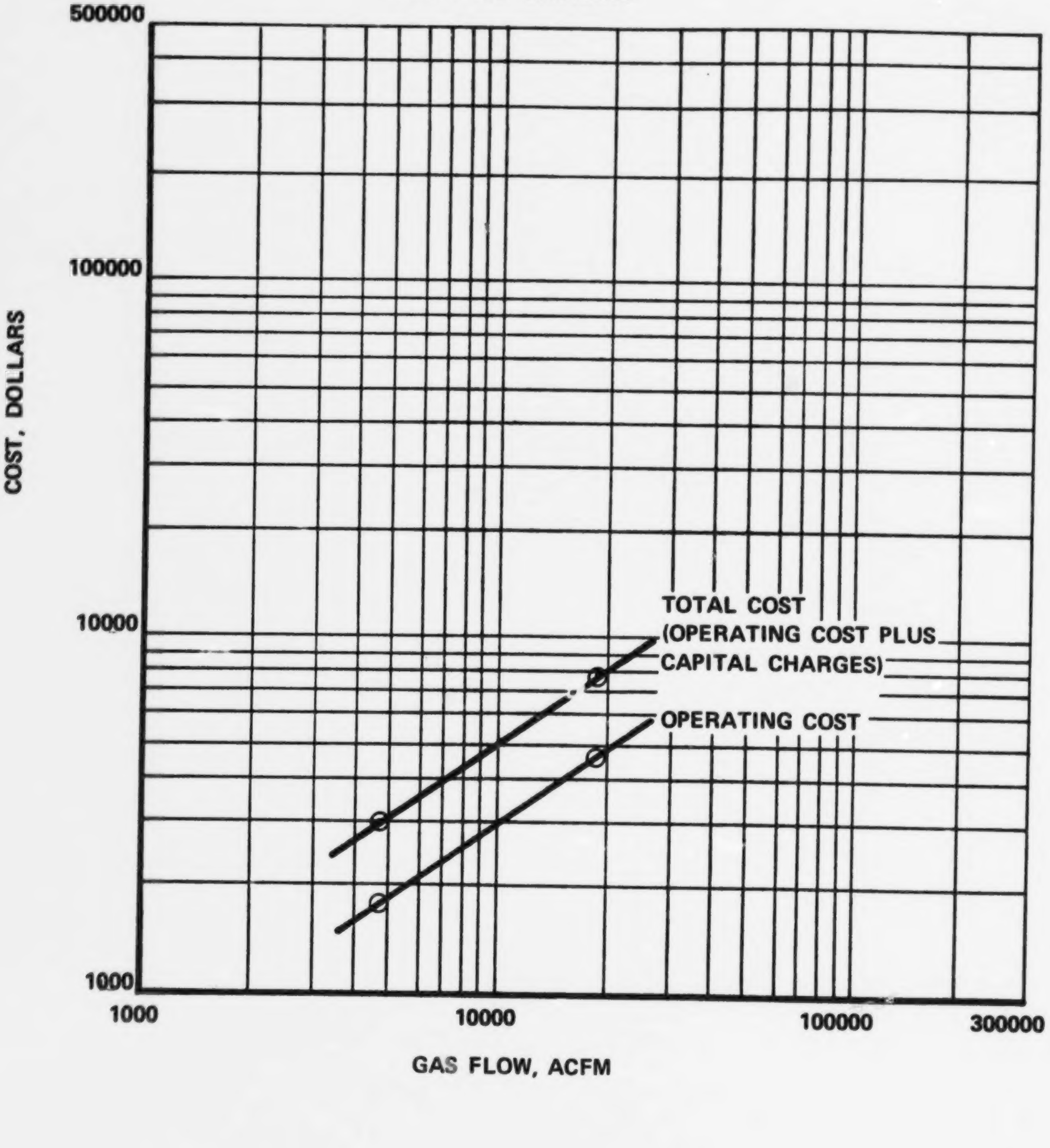


TABLE 79

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF FABRIC FILTERS FOR FEED GRINDING

Population Size – 20   Sample Size – 3

Collector Cost = $5,201

Conf. Level, %	Collector Cost, Dollars Lower Limit	Upper Limit
50	$4,670	$5,733
75	4,171	6,232
90	3,404	6,999
95	2,695	7,708

Collector Cost = $14,395

Conf. Level, %	Collector Cost, Dollars Lower Limit	Upper Limit
50	$12,129	$16,660
75	10,005	18,784
90	6,737	22,053
95	3,715	25,075

FIGURE 59

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF FABRIC FILTERS FOR FEED GRINDING

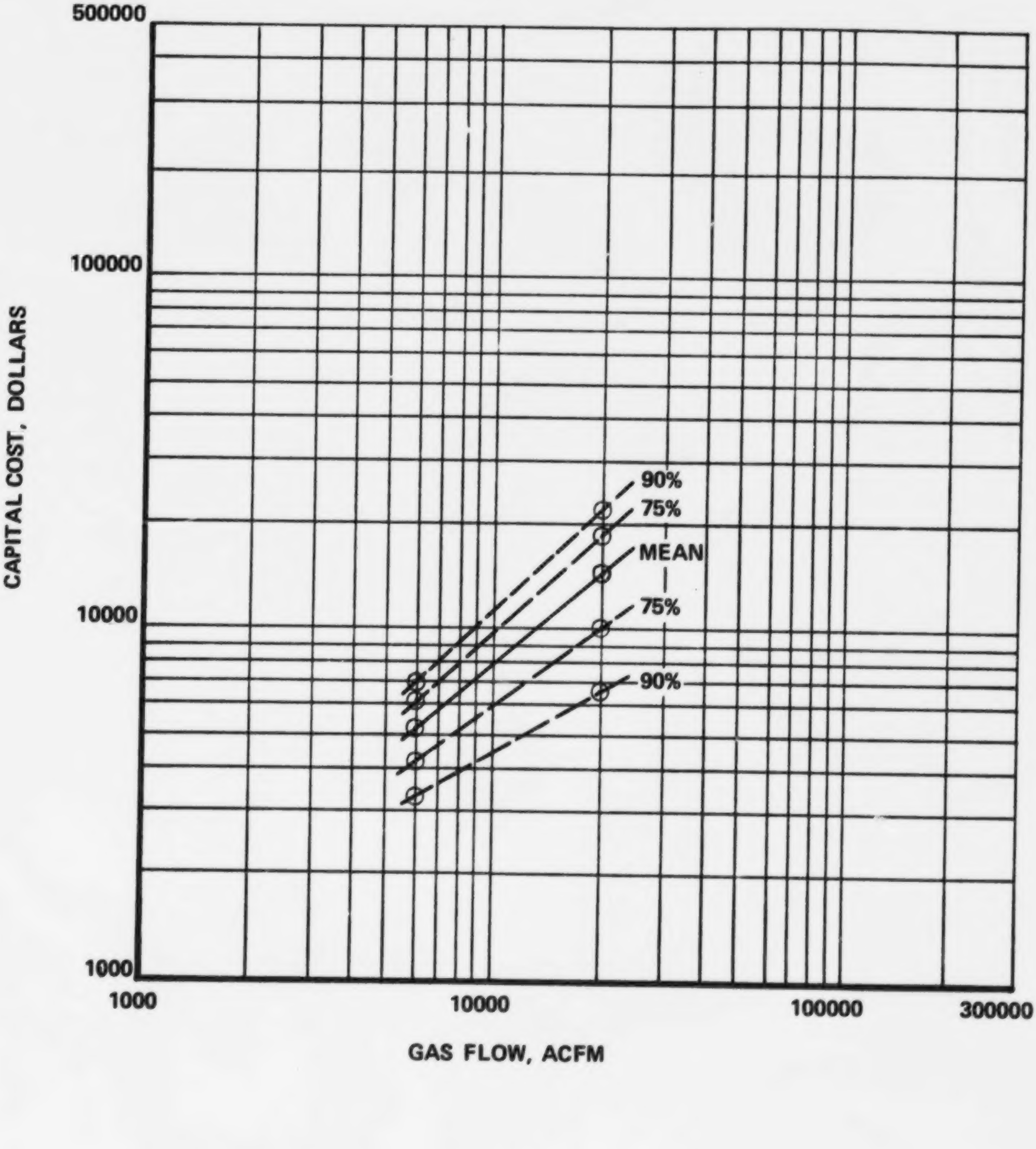


TABLE 80

WET SCRUBBER PROCESS DESCRIPTION FOR FEED FLASH DRYER SPECIFICATION

The scrubber is to deodorize the total gases emitted from the product recovery cyclone of a feed flash drying system. The scrubber will be located outside and will be in use on a continuous basis. Sufficient electric power and fresh water are available at the site. A sewer is available and will accept water in the 4 to 10 pH range, if it contains less than 1 wt.% solids content.

The scrubbing liquor is to consist of a 3 wt.% solution of potassium permanganate buffered to 9.0 pH with borax. Bids should include the following:

1. *Low energy wet scrubber and mist eliminator.*
2. *Necessary fans and motors. Fans should operate at less than 2,000 rpm.*
3. *30 ft stack.*
4. *Recirculation tank.*
5. *Permanganate makeup facilities and storage tank.*
6. *Interconnecting ductwork for all equipment furnished.*
7. *Appropriate control system.*
8. *Necessary provisions for periodic cleaning of manganese dioxide residue.*
9. *Winterizing protection down to -20°F.*

All of the above, except the scrubber proper, should be treated as auxiliaries.

The scrubber must be designed to abate the odor from the effluent stream. However, an estimate of particulate removal efficiency is required.

Each bidder will submit four separate and independent quotations; one for each of two efficiency levels at each of two plant sizes.

TABLE 81

WET SCRUBBER OPERATING CONDITIONS FOR FEED FLASH DRYER SPECIFICATION

	Small	*Large*
Product Rate, lb/hr	*7,360*	*51,300*
Dryer Capacity, lb/hr	*10,500*	*73,300*
Dryer System Discharge Conditions		
Gas Flow, ACFM	*10,000*	*70,000*
Volume % Air	*75*	*75*
Water Content, Vol.%	*10*	*10*
Temperature, °F	*170**	*170**
Gas Flow, SCFM	*8,250*	*57,800*
Organic Content, gr/SCF	*0.2*	*0.2*
Organic Content, lb/hr	*14.2*	*99*
Particulate Loading, gr/DSCF	*0.2*	*0.2*
Odor Concentration, O.U./SCF	*1,500*	*1,500*
Odor Emission Rate, O.U./min.	12.4×10^6	86.7×10^6
Medium Efficiency Case		
Concentration @ Scrubber Outlet O.U./SCF	*500*	*500*
% Odor Removal	*67*	*67*
High Efficiency Case		
Concentration @ Scrubber Outlet O.U./SCF	*100*	*100*
% Odor Removal	*93.2*	*93.2*

**Possibly as high as 250°F*

TABLE 82

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR WET SCRUBBERS FOR FEED FLASH DRYERS

	Medium Efficiency		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	10,000	70,000	10,000	70,000
°F	170	170	170	170
SCFM	8,250	57,800	8,250	57,800
Moisture Content, Vol. %	10	10	10	10
Effluent Contaminant Loading				
gr/ACF	0.2	0.2	0.2	0.2
lb/hr	14.2	99.0	14.2	99.0
Cleaned Gas Flow				
ACFM	9,230	64,570	9,230	64,570
°F	120	120	120	120
SCFM	8,435	59,000	8,435	59,000
Moisture Content, Vol. %	12.5	12.5	12.5	12.5
Cleaned Gas Dust Loading				
Odor Conc., o.u./SCF	500	500	100	100
Odor Rate, o.u./min.	4.22×10^6	29.5×10^6	0.84×10^6	5.9×10^6
Cleaning Efficiency, % *	67	67	93.2	93.2
(1) Gas Cleaning Device Cost (Incl. Fan & Motor)	10,098	33,050	10,478	35,043
(2) Auxiliaries Cost				
(a) Pump(s)	717	1,295	717	1,295
(b) Damper(s)	70	141	70	141
(c) Conditioning, Equipment				
(d) Dust Disposal Equipment	2,167	5,500	2,167	5,500
(3) Installation Cost				
(a) Engineering	16,383	29,577	16,383	29,577
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost	29,435	69,563	29,815	71,556

*Estimated particulate removal for both cases is 99.7%.

FIGURE 60

CAPITAL COSTS FOR WET SCRUBBERS FOR FEED FLASH DRYERS (HIGH EFFICIENCY)

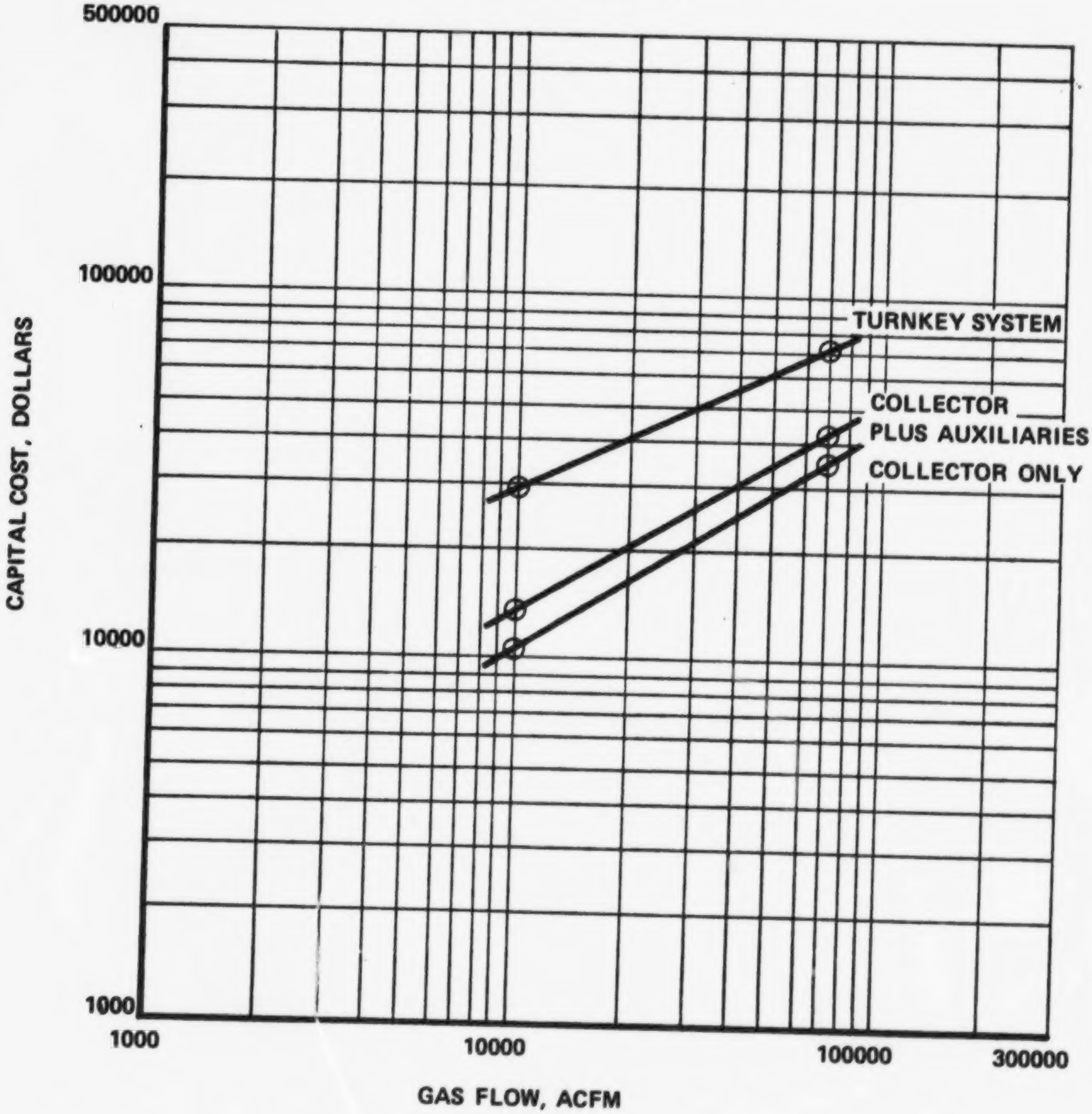


TABLE 83

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR WET SCRUBBERS FOR FEED FLASH DRYERS

Operating Cost Item	Unit Cost	Medium Efficiency		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,000				
Operating Labor (if any)					
Operator					
Supervisor					
Total Operating Labor		170	170	170	170
Maintenance					
Labor					
Materials					
Total Maintenance		1,207	1,207	1,207	1,207
Replacement Parts					
Total Replacement Parts		300	550	300	550
Utilities					
Electric Power	$.011/kw-hr	2,126	14,784	2,196	15,616
Fuel	-	-------	---------	-------	---------
Water (Process)	$.25/M gal	1,800	15,600	1,800	15,600
Water (Cooling)	-	-------	---------	-------	---------
Chemicals, Specify, $KMnO_4$	$0.39/lb	203,830	1,421,753	283,577	1,977,737
Borax	$0.0625/lb	175,090	1,221,286	243,593	1,698,876
Total Utilities		382,846	2,673,423	531,166	3,707,829
Total Direct Cost		384,523	2,675,350	532,843	3,709,756
Annualized Capital Charges		2,944	6,956	2,982	7,156
Total Annual Cost		387,467	2,682,306	535,825	3,716,912

FIGURE 61

ANNUAL COSTS FOR WET SCRUBBERS FOR FEED FLASH DRYERS (HIGH EFFICIENCY)

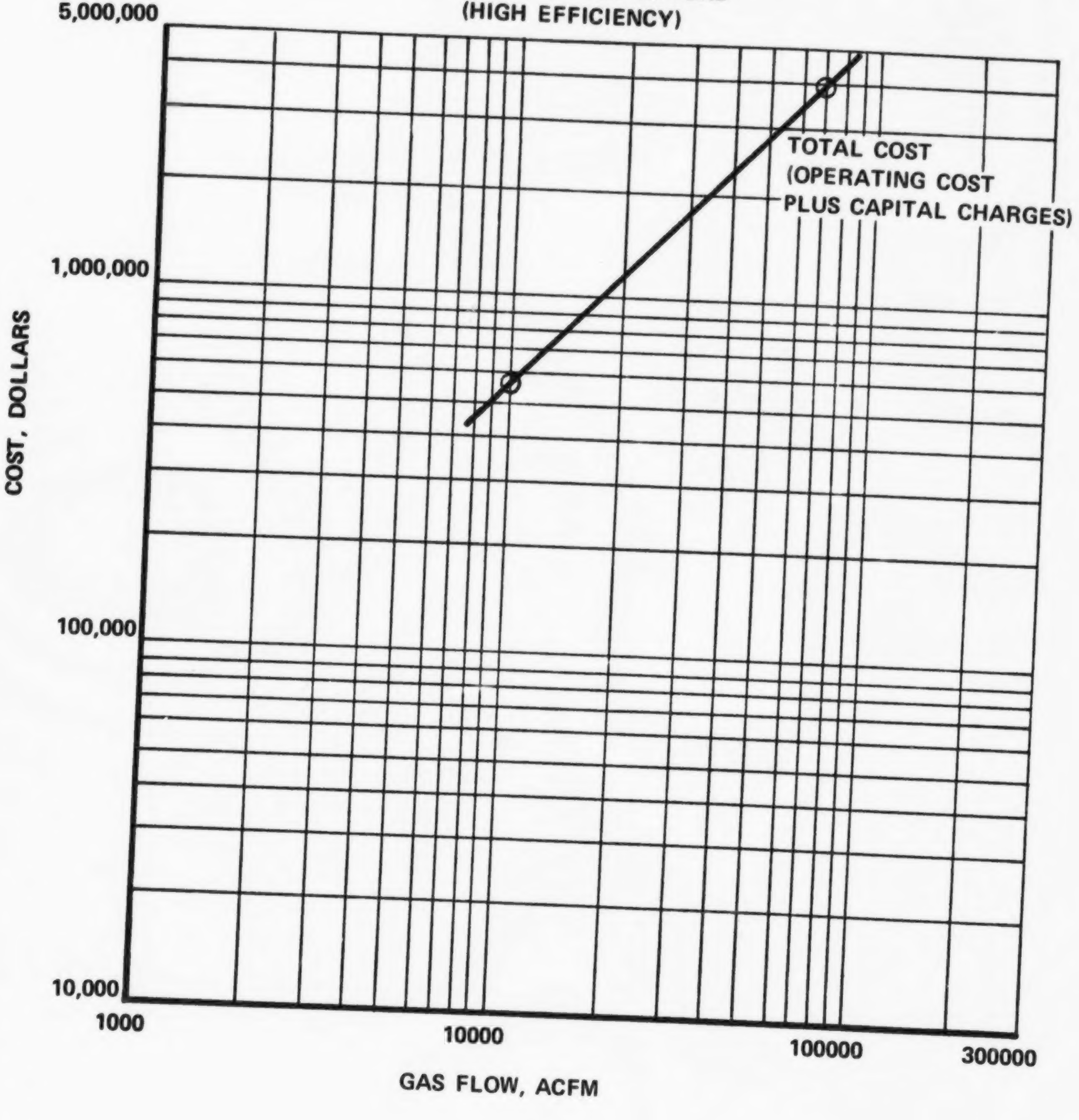


TABLE 84

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR FEED FLASH DRYERS (HIGH EFFICIENCY)

Population Size – 20 Sample Size – 3

Capital Cost = $29,815

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$23,569	$36,061
75	17,713	41,917
90	8,702	50,928
95	370	59,260

Capital Cost = $71,556

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$60,701	$82,412
75	50,523	92,590
90	34,862	108,251
95	20,380	122,733

FIGURE 62

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR FEED FLASH DRYERS (HIGH EFFICIENCY)

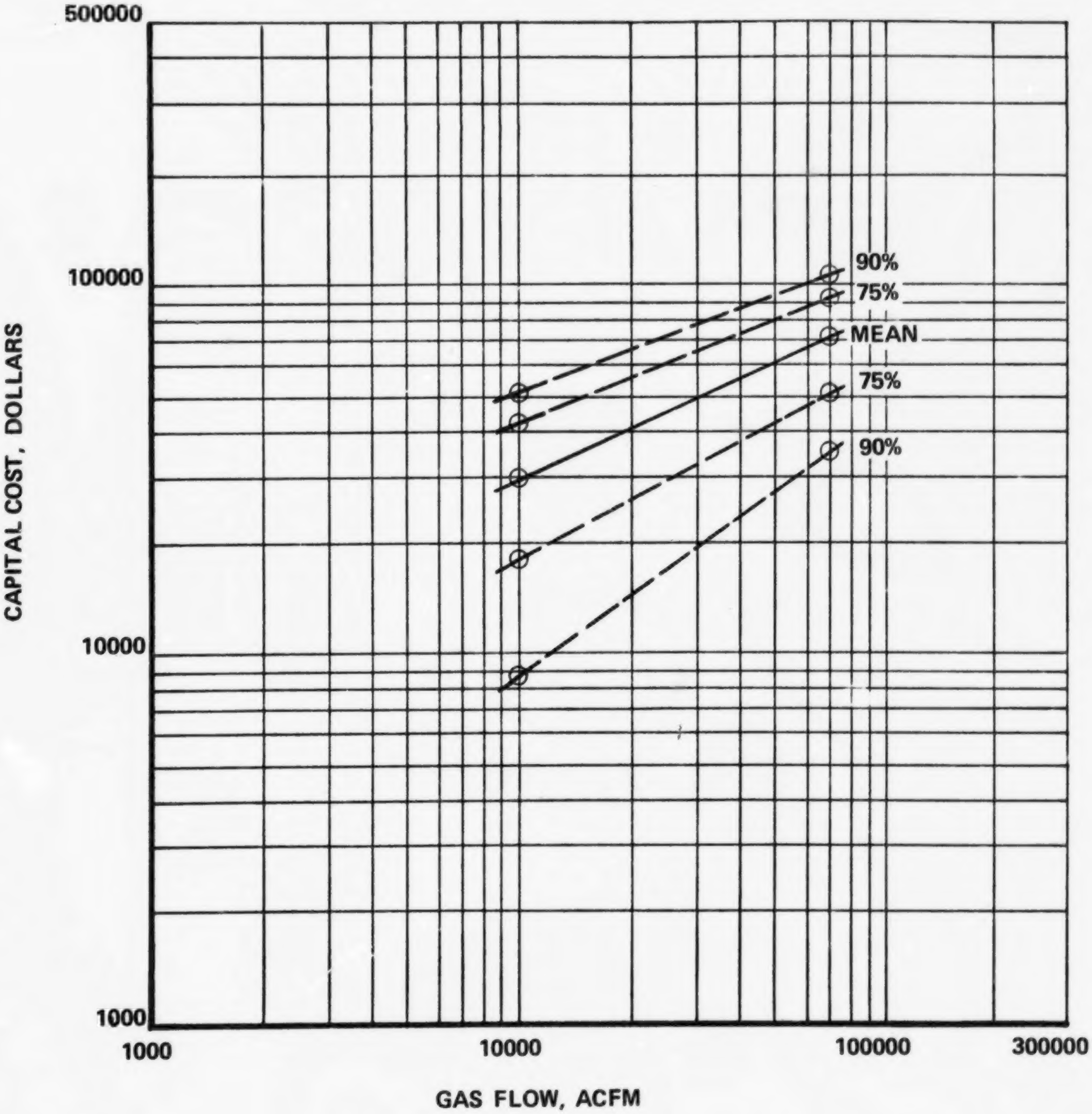


TABLE 85

THERMAL INCINERATOR PROCESS DESCRIPTION FOR FEED FLASH DRYER SPECIFICATION

The incinerator is to deodorize the total gases emitted from the product recovery cyclone of a feed flash drying system. The incinerator will be in use on a continuous basis.

The incinerator is to be natural gas fired. The burner shall be of the 100% secondary air type, utilizing oxygen in the flash drying system effluent for combustion. The burner shall be equipped with a continuous pilot, and shall be controlled to maintain an outlet temperature no higher than 1,500°F. Gas piping, flame failure controls, etc., shall be designed to meet F.I.A. safety standards.*

The incinerator must be designed to abate the odor from the effluent stream. Also, reuse of heat is a prime concern and is to be accomplished by a self-recuperative heat exchanger. Dirty gas will be on the tube side of the heat exchanger.

The incinerator will be located outdoors near the flash dryer outlet. The incinerator shall be maintained under a slightly positive pressure by virtue of a fan at the inlet of the heat exchanger on the odorous gas side. This fan is to be selected to overcome the pressure drop of the incinerator, both sides of the heat exchanger, and the 30 feet of interconnecting ductwork between the product recovery cyclone outlet and the heat exchanger inlet. The fan will be provided with a draft breaker to prevent upsetting the pressure balance at the product recovery cyclone.

**Factory Insurance Association*

TABLE 86

THERMAL INCINERATOR OPERATING CONDITIONS FOR FEED FLASH DRYER SPECIFICATION

One incinerator should be quoted for each size dryer listed below.

	Small	*Large*
Product Rate, lb/hr	*7,360*	*51,300*
Dryer Capacity, lb/hr	*10,500*	*73,300*
Dryer System Discharge Conditions		
Gas Flow, ACFM	*10,000*	*70,000*
Volume % Air	*75*	*75*
Water Content, vol.%	*10*	*10*
Temperature, °F	*170*	*170*
Gas Flow, SCFM	*8,250*	*57,800*
Organic Content, Btu/SCF	*0.5*	*0.5*
Organic Content, gr/SCF	*0.2*	*0.2*
Organic Content, lb/hr	*14.2*	*99*
Particulate Loading, gr/SCF	*0.2*	*0.2*
Odor Concentration, O.U./SCF	*1,500*	*1,500*
Odor Emission Rate, O.U./min	12.4×10^6	86.7×10^6
Incinerator Discharge Conditions		
Gas Flow, SCFM	*≈8,400*	*≈58,700*
Temperature, °F	*1,500*	*1,500*
Odor Concentration, O.U./SCF	*100*	*100*
Odor Emission Rate, O.U./min	$.84 \times 10^6$	5.87×10^6
Combustion efficiency, %	*93.2*	*93.2*
Hot gas discharge from heat exchange, °F	*622*	*622*
Cold gas flow, SCFM	*8,250*	*57,800*
Cold gas temperature, °F	*170*	*170*
Cold gas discharge temperature, °F	*1,058*	*1,058*
Heat exchanger duty, MM Btu/hr	*≈8.2*	*≈8.2*

TABLE 87

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR FEED FLASH DRYERS

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			10,000	70,000
°F			170	170
SCFM			8,250	57,800
Moisture Content, Vol. %			10	10
Effluent Dust Loading				
Odor Conc.,o.u./SCF			1,500	1,500
Odor Rate, o.u./min.			12.4×10^6	86.7×10^6
Organic Content,BTU/SCF			0.5	0.5
Cleaned Gas Flow				
ACFM			31,352	219,240
°F			1,500	1,500
SCFM			8,315	58,195
Moisture Content, Vol. %			11	11
Cleaned Gas Dust Loading				
Odor Conc.,o.u./SCF			100	100
Odor Rate, o.u./min.			0.83×10^6	5.82×10^6
Cleaning Efficiency, %			93.2	93.2
(1) Gas Cleaning Device Cost			59,875	210,700
(2) Auxiliaries Cost				
(a) Fan(s)			3,583	19,205
(b) Pump(s)			-	-
(c) Damper(s)			308	605
(d) Conditioning, Equipment			-	-
(e) Dust Disposal Equipment			-	-
(3) Installation Cost				
(a) Engineering			2,050	4,500
(b) Foundations & Support			2,400	8,750
Ductwork			3,100	10,000
Stack			875	3,200
Electrical			875	3,415
Piping			250	750
Insulation			750	2,000
Painting			50	100
Supervision			700	2,000
Startup			450	1,100
Performance Test			1,750	1,750
Other (Unit Erection)			3,000	6,700
(4) Total Cost			80,016	274,775

FIGURE 63

CAPITAL COSTS FOR THERMAL INCINERATORS FOR FEED FLASH DRYERS

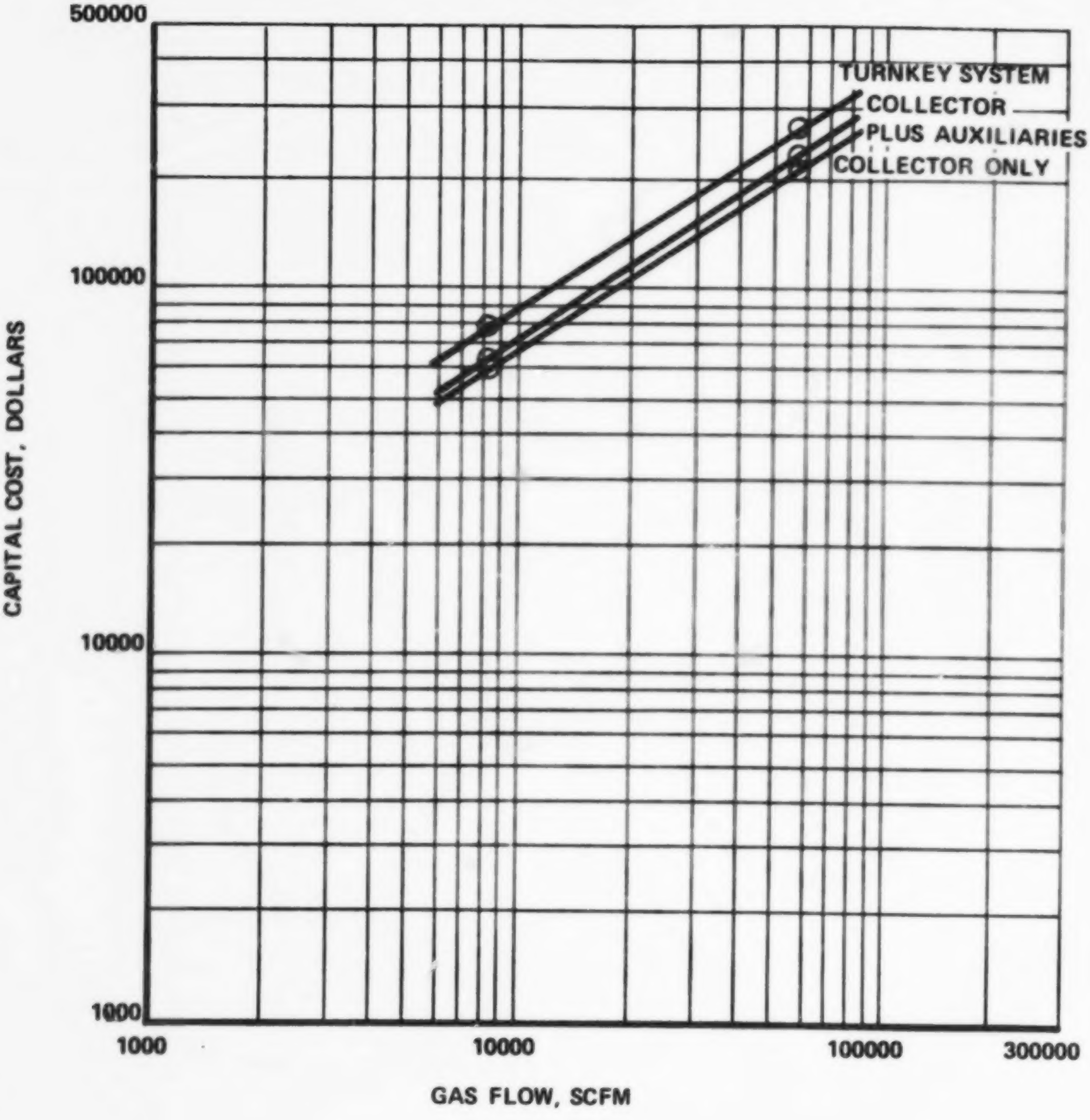


TABLE 88

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR THERMAL INCINERATORS FOR FEED FLASH DRYERS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,000				
Operating Labor (if any)					
Operator	$6/hr.			1,575	2,325
Supervisor				-	-
Total Operating Labor				1,575	2,325
Maintenance					
Labor	$6/hr.			360	720
Materials				65	300
Total Maintenance				425	1,020
Replacement Parts					
Total Replacement Parts				125	275
Utilities					
Electric Power	$.011/kw-h			2,845	19,750
Fuel	$.80/MM Btu			44,890	212,512
Water (Process)				-	-
Water (Cooling)				-	-
Chemicals, Specify				-	-
Total Utilities				47,735	232,262
Total Direct Cost				49,860	235,882
Annualized Capital Charges				8,002	27,478
Total Annual Cost				57,862	263,360

FIGURE 64

ANNUAL COSTS FOR THERMAL INCINERATORS FOR FEED FLASH DRYERS

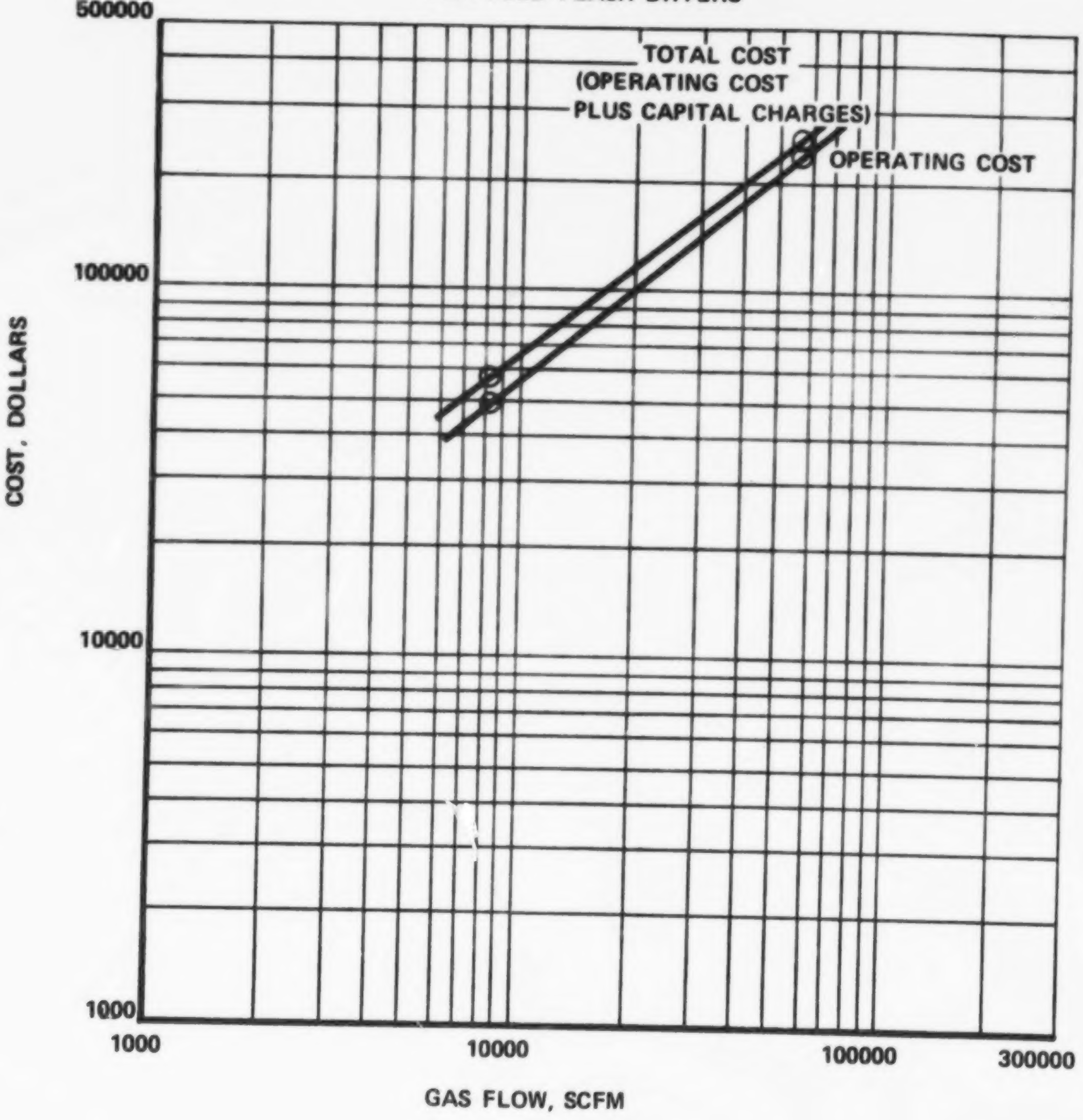


TABLE 89

CONFIDENCE LIMITS FOR CAPITAL COST OF THERMAL INCINERATORS FOR FEED FLASH DRYERS

Population Size – 5 Sample Size – 2

Capital Cost = $80,015

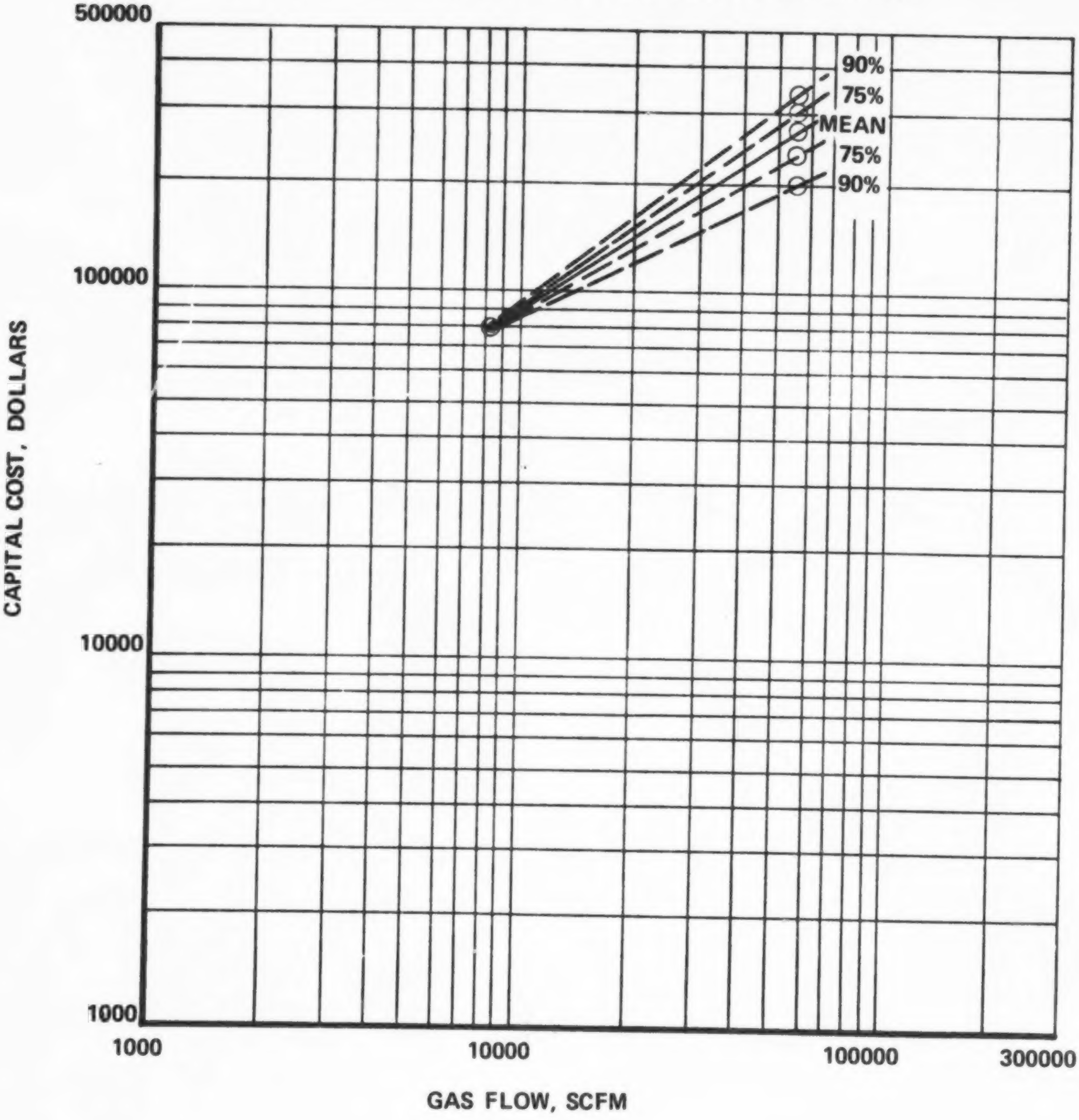
	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$79,891	$80,139
75	79,740	80,289
90	79,446	80,584
95	79,128	80,902

Capital Cost = $274,775

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$259,065	$290,485
75	239,985	309,565
90	202,637	346,913
95	162,298	387,252

FIGURE 65

CONFIDENCE LIMITS FOR CAPITAL COST OF THERMAL INCINERATORS FOR FEED FLASH DRYERS



REFERENCES

1. Matz, Samuel A., Ph.D., Cereal Technology, The AVI Publishing Company, Inc., Westport, Conn., 1970.

2. Air Pollution Engineering Manual, U.S. Dept. of Health, Education, and Welfare, Public Health Services Publication No. 999-AP-40, Cincinnati, Ohio, 1967.

3. "Plant Services Beat Explosion Hazards," Modern Power and Engineering, pp. 40-4, November, 1970.

4. McLouth, Malcolm E. and Paulus, Harold J., "Air Pollution from the Grain Industry," Journal of the Air Pollution Control Association, Vol. 11:7, p. 313-17, July, 1961.

3. PAINT AND VARNISH

The Paint and Varnish industry is one of the oldest manufacturing industries in the United States. The industry is made up of about 1,600 companies operating 1,875 plants.(1) It is well distributed geographically throughout the country and the number of plants or production volume is definitely related to density of population. Even though about 27 companies account for about 57% of the total sales, the industry is one of the few remaining which contains numerous small companies that specialize in a limited product line to be marketed within a geographical region. There are fewer than 20 companies that sell paint nationwide.

The industry is now emerging as a scientific business from its beginning as an art 50 years ago. Even with rapid growth in technology, the industry processing techniques still are not well defined and vary from one producer to another. To add further complication, the industry is technically one of the most complex of the chemical industry. A plant that produces a broad line of products might utilize over 600 different raw materials and purchased intermediates. These materials can be generally classified in the following categories: oils, metallic dryers, resins, pigment extenders, plasticizers, solvents, dyes, bleaching agents, organic monomers for resins and additives of many kinds.

The industry produces an equally large number of finished products which are generally classified as trade sale finishes, maintenance finishes, and industrial finishes.

Trade sale products are stock-type paints generally distributed through wholesale-retail channels and packaged in sizes ranging from 1/2 pint to 1 gallon. A subdivision of trade sale products are maintenance finishes which are used for the protection and upkeep of factories, buildings, and structures such as bridges and storage tanks. Since they are usually stock type, they come under the Department of Commerce definition of trade sales.

The other major type of paint products is industrial finishes which are generally defined as those applied to manufactured products. These finishes such as automotive, aircraft, furniture and electrical are usually specifically formulated for the using industry. Within these major product lines there are literally thousands of different products for many different applications and types of customers. Trade sales finishes and industrial finishes are produced in almost equal volume with the production for this year estimated at 475 million gallons for each type. Trade sales, however, are expected to account for 55% of the dollar sales or about $1,685 million dollars.

PROCESS DESCRIPTION

Starting with all purchased raw material, the manufacturing process for pigmented products is deceptively simple from a process viewpoint. Basically, it consists of mixing or dispersing pigment and vehicle to give the final product. This is schematically illustrated in Figure 66.

The paint vehicle is defined as the liquid portion of the paint and consists of volatile solvent and non-volatile binder such as oils and resins. The non-volatile portion is also called the vehicle solid or film former. The pigment portion of the paint consists of hiding pigments such as titanium dioxide (TiO_2), extenders or fillers such as talc or barium sulfate, and any mineral matter used for flatting or other purposes.

The incorporation of the pigment in the paint vehicle is accomplished by a combination of grinding and dispersion or dispersion alone. When it is necessary to further grind the raw pigment, the pebble or steel ball mills are normally used. With the advent of fine particle grades of pigment and extenders, as well as the wide spread use of wetting agents, the trend is toward milling methods that are based on dispersion without grinding. This dispersion consists of breakup of the pigment clusters and agglomerates, followed by wetting of the individual particles with the binder or vehicle. Some of the more popular methods currently being used are high speed disc impellers, high speed impingement mills and sand mills.

Aside from this dispersion step, pigment paint manufacturing involves handling of raw material as well as handling and packaging of finished product. Operations of a typical plant may be summarized as a raw material and finished product handling problem with a variety of interdispersed batch operations. The interrelationship of all these operations is schematically illustrated in Figure 67. The operations depicted are those of a plant that makes its own resins and produces both trade sale and industrial finishes.

Many of the larger and some of the medium size manufacturers produce a significant amount of their formulation ingredients, primarily resins. A few large manufacturers also produce pigments, modified oils, and basic chemicals. Certain manufacturers produce these ingredients in an amount exceeding their requirements and sell the excess to other manufacturers. A significant number also produce only a portion of their resins and purchase the remainder from their competitors or suppliers who specialize in resin manufacturing.

The manufacturing of resins and varnishes is by far the most complex

FIGURE 66

PAINT MANUFACTURING USING SAND MILL FOR GRINDING OPERATION

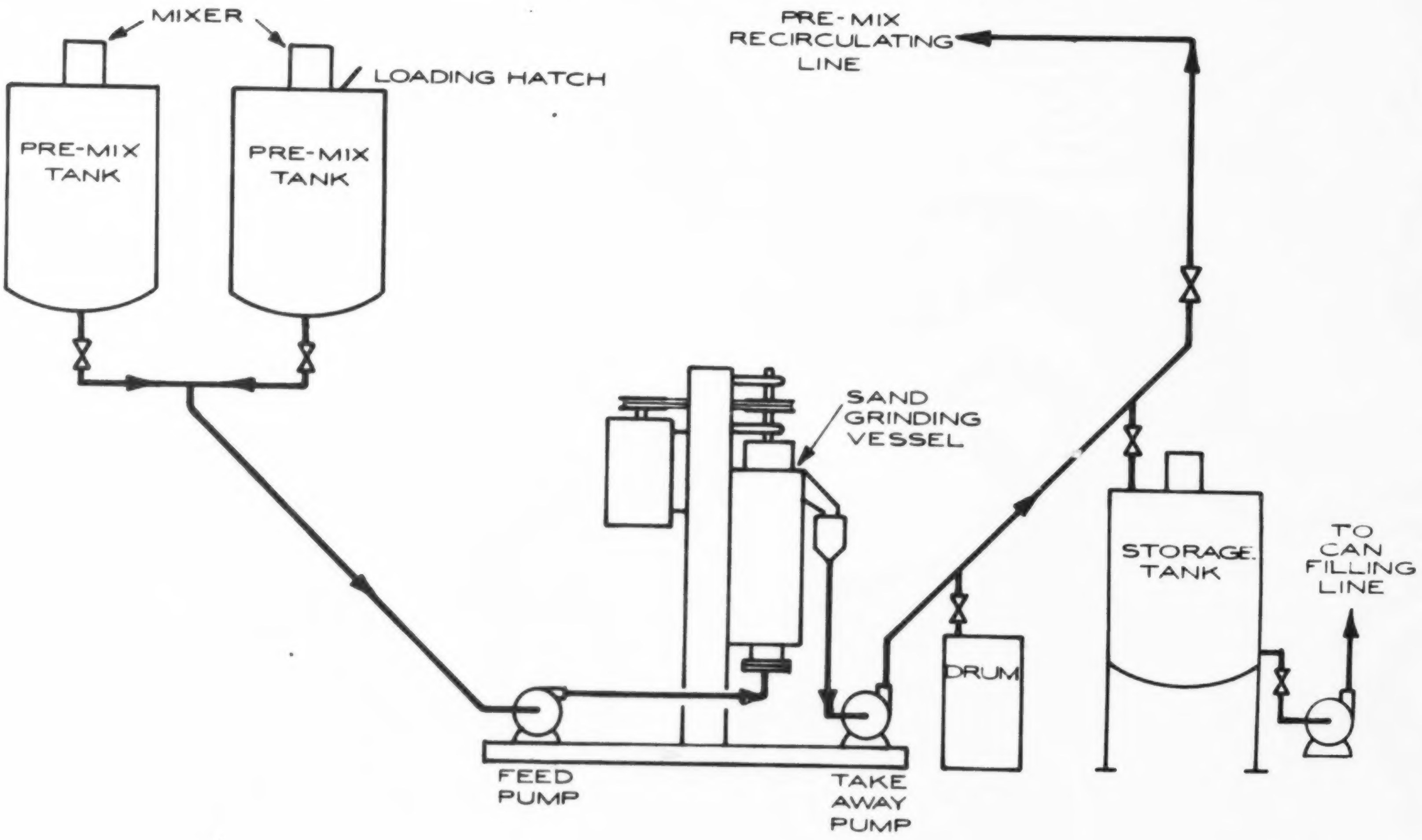
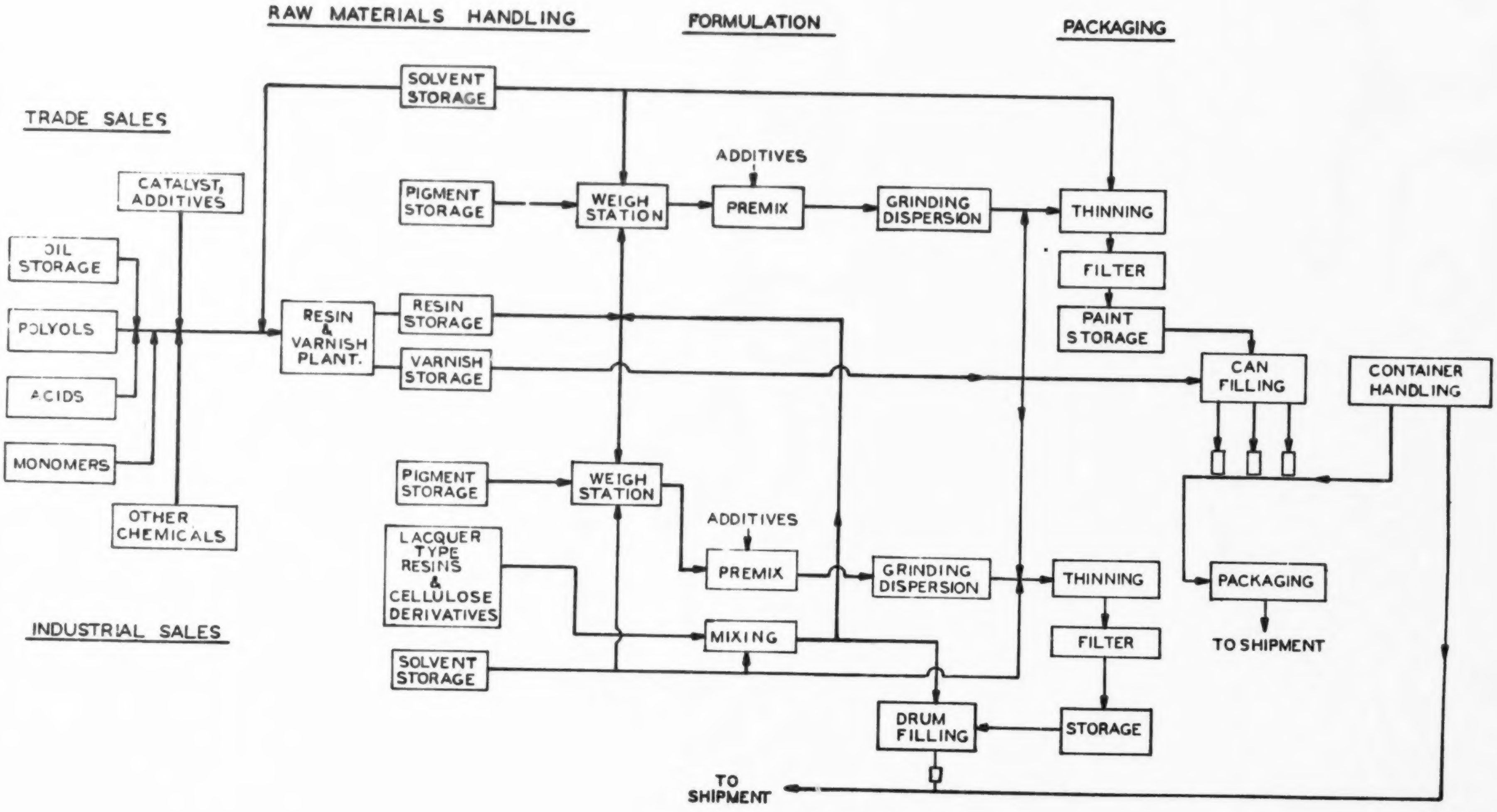

MIXER
LOADING HATCH
PRE-MIX TANK
PRE-MIX TANK
PRE-MIX RECIRCULATING LINE
SAND GRINDING VESSEL
STORAGE TANK
TO CAN FILLING LINE
DRUM
FEED PUMP
TAKE AWAY PUMP

FIGURE 67

MATERIALS FLOW SHEET FOR PAINT MANUFACTURING


RAW MATERIALS HANDLING
FORMULATION
PACKAGING
TRADE SALES
INDUSTRIAL SALES
CATALYST, ADDITIVES
OIL STORAGE
POLYOLS
ACIDS
MONOMERS
OTHER CHEMICALS
SOLVENT STORAGE
PIGMENT STORAGE
RESIN & VARNISH PLANT.
RESIN STORAGE
VARNISH STORAGE
WEIGH STATION
ADDITIVES
PREMIX
GRINDING DISPERSION
THINNING
FILTER
PAINT STORAGE
CAN FILLING
CONTAINER HANDLING
PIGMENT STORAGE
WEIGH STATION
ADDITIVES
PREMIX
LACQUER TYPE RESINS & CELLULOSE DERIVATIVES
MIXING
SOLVENT STORAGE
GRINDING DISPERSION
THINNING
FILTER
STORAGE
DRUM FILLING
PACKAGING
TO SHIPMENT
TO SHIPMENT

process in a paint plant, primarily as the result of the large variety of different raw materials, products and cooking formulas utilized. The complexity begins with the nomenclature used in classification of the final product. Originally, varnishes were all made from naturally occurring material and they were easily defined as a homogeneous solution of drying oils and resins in organic solvents. As new synthetic resins were developed, the resulting varnishes were classified as resins rather than varnishes. Examples of this are alkyd, epoxy, and polyurethane resin varnishes.

There are two basic types of varnishes, spirit varnishes and oleoresinous varnishes. Spirit varnishes are formed by dissolving a resin in a solvent and they dry by solvent evaporation. Shellac is a good example of a spirit varnish. Another material that might fall in this category is lacquer. Technically, lacquers are defined as a colloidal dispersion or solution of nitro-cellulose, or of similar film-forming compounds, with resins and plasticizers, in solvent and diluents which dry primarily by solvent evaporation.

Oleoresinous varnishes, as the name implies, are solutions of both oils and resins. These varnishes dry by solvent evaporation and by reaction of the non-volatile liquid portion with oxygen in the air to form a solid film. They are classified as oxygen convertible varnishes and the film formed on drying is insoluble in the original solvent. A summary of the various types of material used in the production of classical varnishes is given in Table 90.(2)

Varnish is cooked in both portable kettles and large reactors. Kettles are still being used, to a limited extent, by the smaller manufacturers, but some continue to be phased out each year. The very old, coke fired, 30 gallon capacity copper kettles are no longer used. The varnish kettles which are used, have capacities of 150 to 375 gallons. These are fabricated on stainless steel, have straight sides and are equipped with three or four-wheel trucks. Heating is done with natural gas or fuel oil for better temperature control. The kettles are fitted with retractable hoods and exhaust pipes, some of which may incorporate solvent condensers. Cooling and thinning are normally done in special rooms. A typical varnish production operation is illustrated in Figure 68.

The manufacturing of oleoresinous varnishes is somewhat more complex than for spirit varnishes. This manufacture consists of the heating or cooking of oil and resins together for the purpose of obtaining compatability of resin and oil and solubility of the mixture in solvent, as well as for development of higher molecular weight molecules or polymers.

The time and temperature of the cook are the operating variables used to

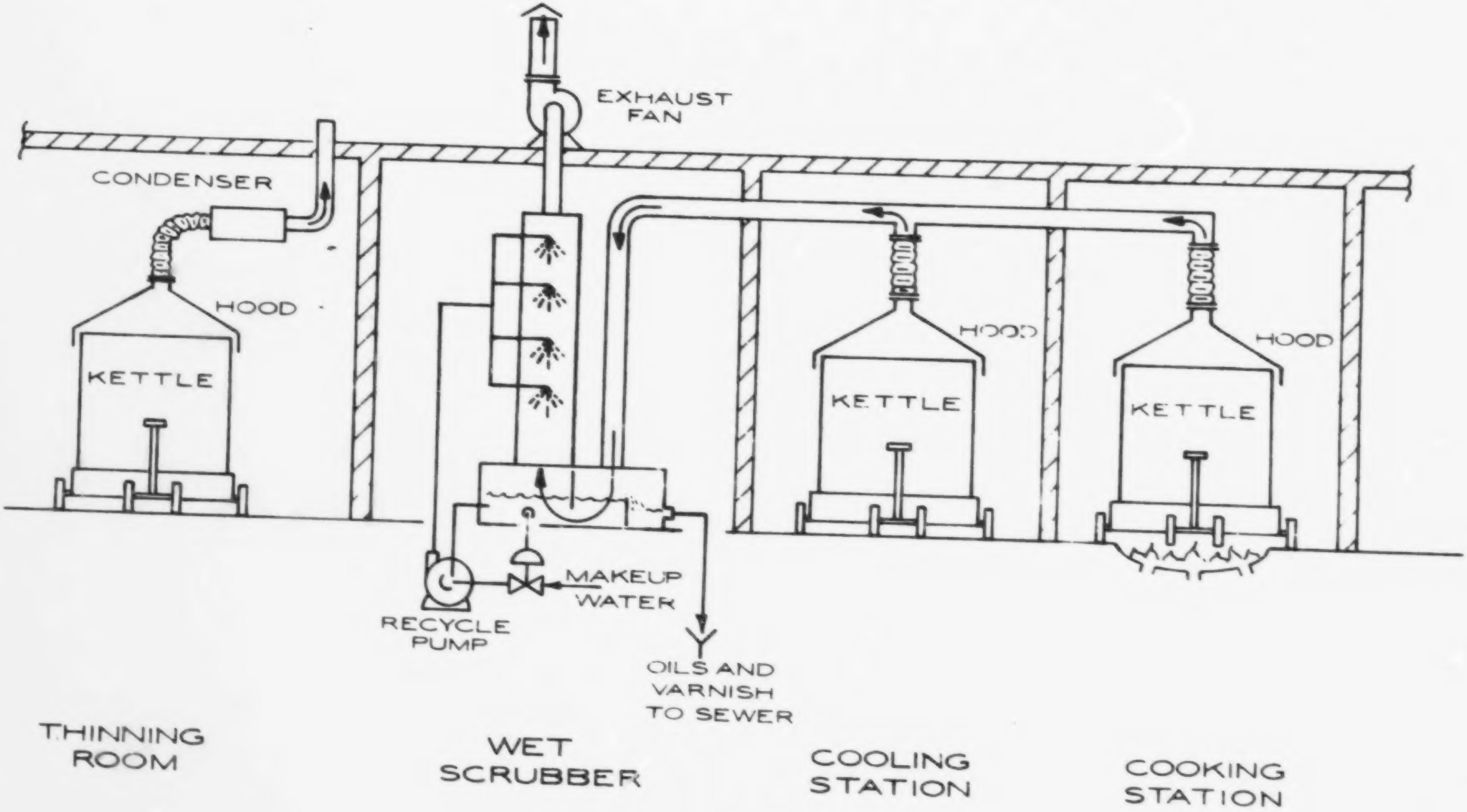
TABLE 90

VARNISH RAW MATERIAL[2]

OILS	RESINS	SOLVENTS	DRYERS	ADDITIVES
1. Hard a. Tung b. Oiticica c. Dehydrated castor oil 2. Soft a. Linseed b. Safflower c. Soya d. Fish e. Tall oil fatty acids 3. Chemically Modified a. Maleic-treatment	1. Natural a. Fossil resin, congo kauri, pontianak b. Semi-fossil resin manila, boea, batu, eastindia c. Recent and crop resin dammar, accroides sandaral mastic, elemi, shellac d. Rosin Gum rosin Wood rosin Tall oil rosin Rosin esters 2. Synthetic Resin a. Maleic resin phenolic resin 3. Metallic Soaps a. Limed rosin b. Zinc resinate	1. Mineral Spirits 2. VM&P Naphtha 3. Toluol 4. Xylol 5. Terpene Solvents	Lead Manganese Cobalt Calcium Zirconium	Anti-skinning Agents Ultra-violet absorber Flattening Agents

FIGURE 68

TYPICAL VARNISH COOKING ROOM


EXHAUST FAN
CONDENSER
HOOD
KETTLE
HOOD
KETTLE
HOOD
KETTLE
RECYCLE PUMP
MAKEUP WATER
OILS AND VARNISH TO SEWER
THINNING ROOM
WET SCRUBBER
COOLING STATION
COOKING STATION

develop the desired end product polymerization or "body". The chemical reactions which occur are not well defined. The resin is a polymer before cooking and may or may not increase in molecular size during the cook. This resin may react with the oil to produce copolymers of oil and resin or it may exist as a homogeneous mixture or solution of oil homopolymers and resin homopolymers.

It is possible to blend resins and heat-bodied oil and obtain the same varnish that can be produced by cooking the resin and the unbodied oils. This indicates that copolymerization is not the fundamental reaction in varnish cooking.

Heat bodying or polymerization of an oil is done to increase its viscosity and is carried out in a kettle in a fashion similar to varnish cooking. The fundamental reaction that occurs is polymerization of the oil monomers to form dimers with a small portion of trimers.

There is a large variety of synthetic resins produced for use in the manufacture of surface coatings. A listing of the more popular resins is given below. They are listed by order of consumption by the coatings industry:(6)

Alkyd
Vinyl
Amino
Epoxy
Acrylic
Styrene Butadiene
Phenolic
Polyester
Urethane
Silicone

By far the most widely used of these resins are the alkyds and the vinyls. Alkyd consumption is approximately five times that of the vinyl, which is approximately twice that of the amino resins. Further discussion will concentrate on alkyd resins.

Alkyd resins comprise a group of synthetic resins which can be described as oil-modified polyester resins. They are produced from the reaction of polyols or polyhydric alcohol, polybasic acid and oil or fatty monobasic acid. A listing and discussion of commonly used raw materials will follow.

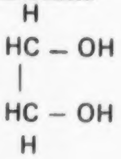
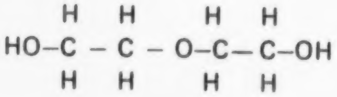
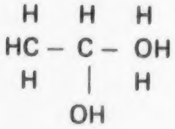
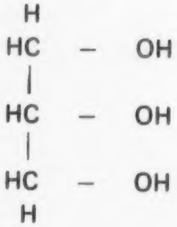
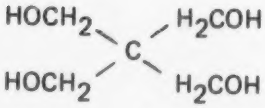
1. Oils or Fatty Acid(2)

   - Linseed
   - Soybean
   - Safflower
   - Tall Oil Fatty
   - Tall Oil
   - Fish
   - Tung (minor)
   - Oiticica (minor)
   - Dehydrated Castor (minor)

   - Castor
   - Coconut
   - Cottonseed
   - Lauric Acid
   - Pelargonic Acid
   - Isodecanoic Acid

The materials in the first column are oxidizing or drying types. The materials in the second column are non-oxidizing and yield soft non-drying alkyds which are used primarily as plasticizers for hard film resins. The acids shown in this column are the only materials that are strictly synthetic in origin.

2. Polyols

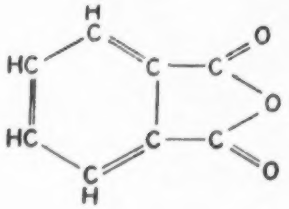
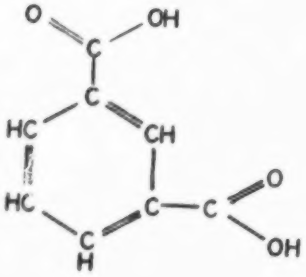
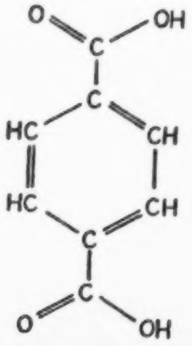
Name	Formula	Form
Ethylene glycol	 $\begin{array}{c} H \\ HC - OH \\ \vert \\ HC - OH \\ H \end{array}$	Liquid
Diethylene glycol	 $\begin{array}{ccccccccc} & H & & H & & H & & H & \\ HO- & C & - & C & - O - & C & - & C & -OH \\ & H & & H & & H & & H & \end{array}$	Liquid
Propylene glycol	 $\begin{array}{ccccc} H & & H & & H \\ HC & - & C & - & OH \\ H & & \vert & & H \\ & & OH & & \end{array}$	Liquid
Glycerine CP-95% glycerine Super-98%	 $\begin{array}{ccc} H & & \\ HC & - & OH \\ \vert & & \\ HC & - & OH \\ \vert & & \\ HC & - & OH \\ H & & \end{array}$	Liquid
Pentaerythritol		White Solid

Glycerol or glycerine was the first polyol used for alkyds. It is also the most widely used polyol for alkyds.

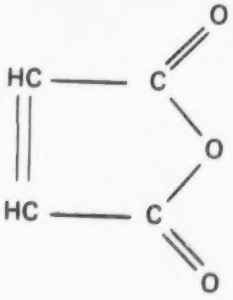
The second polyol, based on usage, is pentaerythritol (PE), which came into common use in the 1940's. PE is supplied as "technical grade" material and contains mono, di, tri, and polypentaerythritol. The material consists primarily of the mono form which was illustrated previously in the list of polyols.

The important distinguishing feature of the various polyols is the number of potentially reactive hydroxyl groups in the molecule, known as functionality. The glycols with a functionality of two produce only straight chain polymers and their resins are soft and flexible. The resultant products are used primarily as plasticizers for hard resins. Glycerine has a functionality of three and is used primarily in short and medium oil alkyds. Pentaerythritol, with a functionality of four, cross links to a greater extent, forming harder polymers. It is ideal for use in long oil alkyds.

3. Acids and Anhydrides

Name	Formula	Form
Phthalic anhydride (ortho)		White solid
Isophthalic acid (meta)		White needles
Terephthalic acid (para)		White crystals

Maleic anhydride



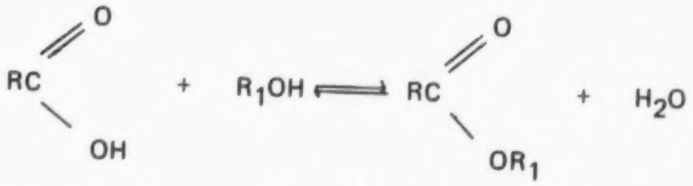
White solid

The acidic material can be used as an acid or anhydride. The anhydride is formed from two molecules of acid minus a molecule of water or removal of one molecule of water from a di-acid. It is preferred, since it reacts faster and yields less water for removal from the cook.

For many years, ortho phthalic anhydride (PA) was the only polybasic acid used in substantial proportions in alkyds. It still remains the predominant dibasic acid. PA is produced from the catalytic oxidation of naphthalene or ortho-xylene.

The chemistry of alkyd resin systems is very complex. So much so that theoretical considerations offer only a good starting point. Final formulae and variations are developed by trial and error changes, based on performance requirements and shortcomings of previous batches.

Condensation is the reaction basic to all polyester resins, including alkyds. This reaction follows the elementary equation for esterification as shown below:



Acid + Alcohol Ester + Water

For Alkyd Resins

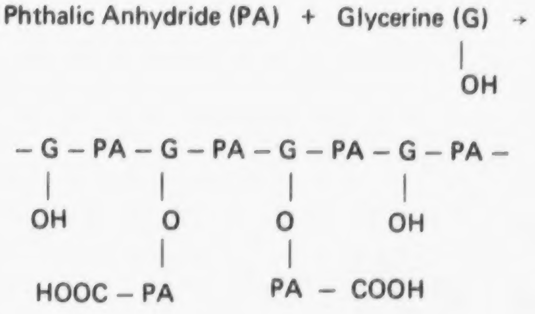
$$\text{PA} + \text{Glycerine} \rightleftharpoons \text{Ester} + H_2O$$

The ester monomer formed is very complex and further reacts to form large polymers called resins. The polymers formed are low in molecular weight by comparison to other resins. For example alkyd resins have molecular weights ranging from 1,000 to 7,000 while some vinyl and acrylic resins have average molecular weights in excess of 100,000 and in some cases as high as 500,000.

The alkyd polymers also react with oil or fatty acid and are generally classified by the amount of oil or PA used in the formulation, as described below:

	% Oil	% PA
Short Oil	33 to 45	35
Medium Oil	46 to 55	30 to 35
Long Oil	56 to 70	20 to 30
Very Long Oil	71 up	20

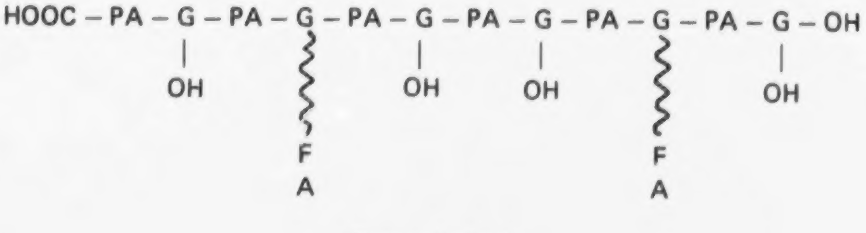
The resulting reactants of the PA, polyol and oil may be represented in part as shown below.



$$\text{Phthalic Anhydride (PA)} + \underset{\substack{|\\ \text{OH}}}{\text{Glycerine (G)}} \rightarrow$$

$$-\underset{\substack{|\\ \text{OH}}}{\text{G}}-\text{PA}-\underset{\substack{|\\ \text{O}\\ |\\ \text{HOOC}-\text{PA}}}{\text{G}}-\text{PA}-\underset{\substack{|\\ \text{O}\\ |\\ \text{PA}-\text{COOH}}}{\text{G}}-\text{PA}-\underset{\substack{|\\ \text{OH}}}{\text{G}}-\text{PA}-$$

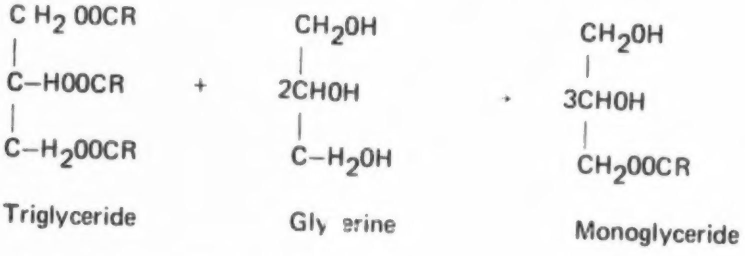
This will then react with the long chain oil monoglyceride or fatty acid (FA) to yield:



$$\text{HOOC}-\text{PA}-\underset{\substack{|\\ \text{OH}}}{\text{G}}-\text{PA}-\underset{\substack{\wr\\ \text{F}\\ \text{A}}}{\text{G}}-\text{PA}-\underset{\substack{|\\ \text{OH}}}{\text{G}}-\text{PA}-\underset{\substack{|\\ \text{OH}}}{\text{G}}-\text{PA}-\underset{\substack{\wr\\ \text{F}\\ \text{A}}}{\text{G}}-\text{PA}-\underset{\substack{|\\ \text{OH}}}{\text{G}}-\text{OH}$$

Short oil alkyd

Alkyds can be manufactured directly from a fatty acid, polyol and acid or from the fatty acid oil, polyol and acid. The second combination (oil, glycerine and PA) produces glyceryl phthalate which is insoluble in the oil and precipitates. This problem can be overcome by first converting the oil to a monoglyceride by heating with a polyol in the presence of a catalyst. This process is called alcoholysis of the oil. The basic reaction is shown below:



$$\begin{array}{l} CH_2OOCR \\ | \\ C{-}HOOCR \\ | \\ C{-}H_2OOCR \end{array} + \begin{array}{l} CH_2OH \\ | \\ 2CHOH \\ | \\ C{-}H_2OH \end{array} \rightarrow \begin{array}{l} CH_2OH \\ | \\ 3CHOH \\ | \\ CH_2OOCR \end{array}$$

This is an ester interchange reaction with no loss of water.

When fatty acid rather than oil is used as the starting material, this is called the "one-stage" process. In this process, the fatty acid and glycerine are added to the kettle, the agitator is started and heat is introduced. When the batch reaches 440° F, the PA is slowly added and cooking continued for another 3 to 4 hours until the desired body and acid number are reached.

If the fusion process is being used, a continuous purge of inert gas is maintained to remove the water formed in the reaction. This water may also be removed by what is known as the solvent process. It is similar to the fusion process except that about 10% aromatic solvent (usually xylene) is added at the start. The vaporized solvent is passed into a condenser. The condensate then flows to a decant receiver for separation of reaction water. Recovered solvent is returned to the reactor.

As discussed earlier, when oil is used rather than fatty acid, the alkyds are produced in a two stage process. In the first stage the monoglyceride is first produced from the linseed oil and glycerol. Catalyst and oil are added and the alcoholysis of the polyol and oil is carried out between 450 and 500° F until the desired end point is reached. When the alcoholysis is completed, any additional polyol needed is added.

Following this, the required amount of PA and esterification catalyst are slowly added. If solvent cooking is to be used, the solvent is also added at this time. Cooking then proceeds as before.

A typical manufacturing formula for a 50% oil-modified glyceryl phthalate alkyd using the two stage process is given below.(4)

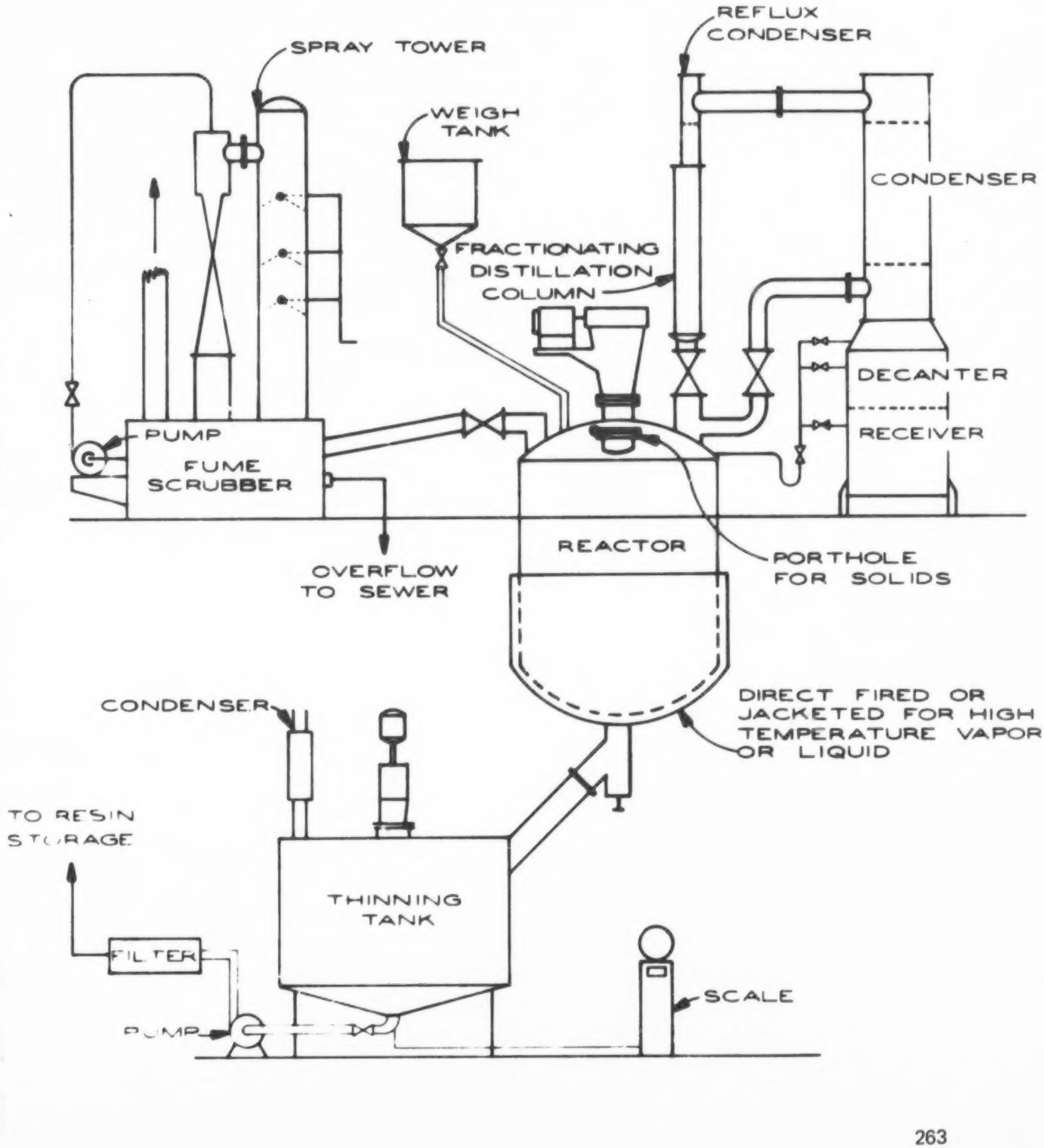
	lb
First stage	
Linseed oil	51.3
Glycerol (95%)	12.8
Catalyst, $Ca(OH)_2$	0.026
Second stage	
Glycerol (95%)	6.2
Phthalic Anhydride	39.7
Catalyst	
Methyl p-Toluene Sulfonate	0.2
	110.2
Approx. Loss	10.2
Solids Yield	100.0

Alkyd and other resins are cooked in closed set kettles more properly called reactors. They vary in size in commercial production from 500 to 10,000 gallons. A typical reactor system is shown in Figure 69. They are generally fabricated of Type 304 or 316 stainless steel with well polished surfaces to assure easy cleaning. Design pressure is usually 50 psig. These reactors may be heated electrically, direct fired with gas or oil, or indirectly heated using a heat transfer medium such as Dowtherm$^{(R)}$. They are also equipped with a manway, sight-glass, charging and sampling line, condenser system, weigh tanks, temperature measuring devices and agitator. The manway is used both for charging solid material and for access to the kettle for cleaning and repair.

The reactor may be equipped with a variety of different condenser systems. The system shown in Figure 69 includes a packed fractionating column, a reflux condenser and a main condenser. The condensers are water cooled shell and tube type and may be either horizontally or vertically inclined. Vapors are processed and condensed on the tube side and drain to a decant receiver for separation and possible return as solvent to the reactor. A dual function aspirator Venturi scrubber is often added to the system. It is used to ventilate the kettle during addition of solid materials and may also remove entrained unreacted or vaporized solids and liquids from the venting gases.

FIGURE 69

MODERN RESIN PRODUCTION SYSTEM


REFLUX CONDENSER
SPRAY TOWER
WEIGH TANK
CONDENSER
FRACTIONATING DISTILLATION COLUMN
DECANTER
RECEIVER
PUMP
FUME SCRUBBER
REACTOR
PORTHOLE FOR SOLIDS
OVERFLOW TO SEWER
DIRECT FIRED OR JACKETED FOR HIGH TEMPERATURE VAPOR OR LIQUID
CONDENSER
TO RESIN STORAGE
THINNING TANK
FILTER
SCALE
PUMP

Thinning tanks are always included as part of the reactor system. They are normally water cooled and equipped with a condenser and agitator. The partially cooled finished alkyd is transferred from the reactor to the partially filled thinning tank. Since most alkyd resins are thinned to 50% solids, the capacities of these tanks are normally twice the capacity of the reactors. These tanks are also frequently mounted on scales so that thinning solvents may be accurately added.

The final step in a reactor system is filtering of the thinned resin prior to final storage. This is normally done while it is still hot. Filter presses are the most commonly used filtering device.

The manufacturing procedures and equipment used for the production of other resins listed at the beginning of this discussion are quite similar. The major differences are the raw materials and the process steps utilized. A detailed discussion of these other resins is beyond the scope of this narrative.

NATURE OF GASEOUS DISCHARGE

There are two major types of emissions from a paint plant. These are non-fugitive and fugitive. Non-fugitive emissions are those that are collected by and confined within an exhaust system. Fugitive emissions are those that escape into the plant atmosphere from various operations and exit the plant building through the doors and windows in an unregulated fashion.

In today's typical paint plant there are two types of fugitive emissions. These are pigment particulate and paint solvents. In a small percentage of the plants an attempt is made to collect these emissions. The incentive for doing so is based on insurance requirements as well as occupational health and safety rather than air pollution considerations or regulations. The newly passed OSHA regulations will have a dramatic effect on the paint industry practice and necessitate the regulation of fugitive emissions.

PARTICULATE CONTAMINANTS

Fugitive particulate emissions consist primarily of the various pigments used. As a general rule, the pigments are received and stored in 25 to 50 pound paper sacks or fiber drums. Modern pigment manufacturing has developed fine sized pigment, 0.05 to 0.25 microns, for ease of dispersion into the paint vehicle. Loading of these fine pigments into grinding equipment results in fugitive particulate dust emissions into the surrounding plant areas. This dust is either collected by a ventilation and exhaust system or allowed to settle and later collected as part of the general housekeeping requirements.

A variety of resins are received as granular or flaked solids which are of large size and do not result in a fugitive dust emission. The manufacturer of these solid resins, however, does encounter fugitive emission problems in his flaking or grinding operations.

GASEOUS CONTAMINANTS

Solvent emissions occur in almost every phase of paint and varnish manufacturing and in numerous locations throughout individual plants. A listing of emission points is given below.

	Location	Operation	Temp., °F	Pressure
1.	Resin Plant	Thinning	200 to 300° F	Atmospheric
2.	Resin Plant	Filtering	200 to 300° F	Atmospheric
3.	Resin Plant	Storage Tanks	100° F	Atmospheric
4.	Paint Plant	Blending Tanks	Ambient	Atmospheric
5.	Paint Plant	Milling	Ambient	Atmospheric
6.	Paint Plant	Dispersion	Ambient	Atmospheric
7.	Paint Plant	Holding Tank	Ambient	Atmospheric
8.	Paint Plant	Filtering	Ambient	Atmospheric
9.	Paint Plant	Packaging	Ambient	Atmospheric

The extent of these emissions varies with the type of operation and the effort extended to control atmospheric losses. The high temperature thinning and filtering result in the largest emissions while packaging in drums and cans contributes the smallest emission. Other operations contribute intermediate emissions which vary depending on the degree of control exercised and the vapor pressure of the solvent used.

In some cases, efforts are made to collect fugitive emissions by use of ventilation hoods and a closed exhaust system. More frequently, however, they are exhausted from the building by general building exhaust fans which ventilate areas having the highest contaminant concentration.

Resin plants or paint plants producing resins and varnishes are likely to have a number of regulated emissions. These emissions consist primarily of gaseous hydrocarbons in air or inert gas streams. The three major sources of these non-fugitive emissions are:

1. Varnish cooking
2. Resin cooking
3. Thinning

Other less concentrated streams that may or may not be regulated are:

1. Storage and rundown tank vent systems
2. Filter press vent systems
3. Sandmill vent systems

Considerable effort has been expended to identify the various types of chemical compounds emitted during a varnish cook. The majority of this work was done in the 1950's and is well summarized by R. L. Stenburg(3) in the H.E.W. Technical Report A58-4. A copy of his summary is included here as Table 91.(3) In general, one or more of the following compounds are emitted, depending upon the ingredients in the cook and the cooking temperature: water vapor, fatty acids, glycerine, acrolein, phenols, aldehydes, ketones, terpene oils and terpene. These materials are mainly decomposition products of the varnish ingredients.

Varnishes and oils are cooked or boiled at temperatures from 200 to 650°F. At about 350°F decomposition begins and continues throughout the cooking cycle which normally runs between 8 and 12 hours. The quantity, composition and rate of emissions depend upon the ingredients in the cook as well as the maximum temperature, the length, the method of introducing additive, the degree of stirring and the use of inert gas blowing. In general, the emissions will average between one to three percent of the charge in oil bodying and three to six percent in varnish cooking. Aside from academic interest, the exact chemical structure of these emissions is not too important. Of more importance are the characteristics of the emissions related to ease of removal by the applicable pollution control devices.

Modern resin reactors and varnish cookers account for the majority of clear coatings production in the paint and varnish industry. As described earlier, these products are cooked in larger more carefully controlled reactors equipped with product recovery devices which also help reduce atmospheric emission.

As with the old varnish kettles, the amount of emissions varies with the type of cook, the cooking time, the maximum temperature, the initial

TABLE 91

COMPOSITION OF OIL AND VARNISH FUMES(3)

Bodying Oils	Running Natural Gums	Manufacturing Oleo resinous Varnish	Manufacturing Alkyd Varnish
Water vapor	Water vapor	Water vapor	Water vapor
Fatty acids	Fatty acids	Fatty acids	Fatty acids
Glycerine	Terpenes	Glycerine	Glycerine
Acrolein	Terpene Oils	Acrolein	Phthalic anhydride
Aldehydes	Tar	Phenols	Carbon dioxide
Ketones		Aldehydes	
Carbon dioxide		Ketones	
		Terpene Oils	
		Terpenes	
		Carbon dioxide	

ingredients as well as the type and method of introducing ingredients.

For solvent cooking the quantity of emission does not vary significantly with the size of the reactor but is rather more a function of the volatility of the solvent being used and the size and/or efficiency of the condenser. Since there is no sparge gas used in solvent cooking, exhaust volumes are small and consist primarily of non-condensed solvent fumes. Emissions will run from 0.1 to 0.5 pounds per hour and will be less cyclic in nature than for fusion cooks.

Emissions during fusion cooking run much higher and vary with the size of the reactor. The total exhaust volume is dependent primarily on the sparge rate of inert gas. Dean H. Parker(4) indicated typical sparge rates of 0.04 ft^3/min/gal of charge during the first hour, 0.02 ft^3 during the second, and 0.01 ft^3 during the remainder of the cook. The exhaust rate will average from 2 ft^3/min/100 gallons of capacity on small reactors to 1 ft^3/min/100 gallons of capacity on large reactors. A summary of source test results from a variety of resin reactors is presented in Table 92.(11)

Since fusion cooking is a cyclic batch process, the concentration of emission will vary from the start to finish of the cook. Hydrocarbon concentration will vary from 15,000 to 80,000 ppm as methane equivalent, depending on the time of the cycle and the type of cook. There are at least 100 different emission curves that could be encountered if one tried to cover all of the different cooking formulas. Particulate phthalic anhydride (PA) is also emitted from the kettle and concentration levels vary depending on cycle time, types of cook, method of charging and type of PA used. Charging of liquid PA rather than dry solid PA significantly reduces the emission rate. However, if the linear velocity of the sparge gas is maintained below 150 ft/min, the carry-over of PA is also significantly reduced. Entrained and sublimed PA will run between 1 to 3 pounds per hour over a period of 50 to 70 minutes during and following the charging period. Plots of hydrocarbon emission level vs. time for three of many possible cooks are given as Figures 70, 71 and 72.(10) These emission concentrations are those measured directly out of a closed kettle or reactor.

Figure 72 shows typical variations in emissions from one batch to another when cooking the same product in the same kettle. Variations twice as great as this are not uncommon. Emissions increase dramatically and rapidly as indicated on Figures 70 and 71 whenever the loading hatches are opened. This is a result of forced exhaust of the kettle to prevent spillage of fumes into the room from the open hatch.

TABLE 92

EMISSION DATA SUMMARY

Resin Type	Reactor Size (gal)	EMISSION: RANGE (lb/hr)	EMISSION: AVERAGE (lb/hr)	Major Component	Average Reaction Temp. (° F.)	
Alkyds	4,000	0.07-1.45	0.7	Xylene	410-460	Solvent Cook, based upon 4 resins.
	1,000	0.17-0.75	0.3	Xylene		Solvent/fusion cook, based upon 7 resins.
	1,000	0.44-1.74	1.0	Xylene		Solvent cook, based upon 4 resins.
						Emissions exceeded 8 pounds/hour during second stage changing.
Polyester	4,000	1.32-27.3	12.9	Propylene Glycol	400-410	Fusion cook, based upon 4 resins.
						Emissions exceeded 8 pounds/hour during changing exotherm, and sparging.
	1,000	0.35-2.23	1.18	Propylene Glycol		Fusion cook, based upon 8 resins.
Acrylics, Melamines	4,000	0.06-0.21	0.10	Butanol	220-250	Solvent cook, based upon 11 resins.
	1,000	0.15-0.16	0.15	Butanol		Solvent cook, based upon 2 resins.

FIGURE 70

HYDROCARBON EMISSION FOR DOWTHERM KETTLE (10)

VAPOR TEMPERATURE RANGED FROM 100°F TO 340°F
KETTLE TEMPERATURE FROM 80°F TO 500°F
COOK RAN 20 HOURS

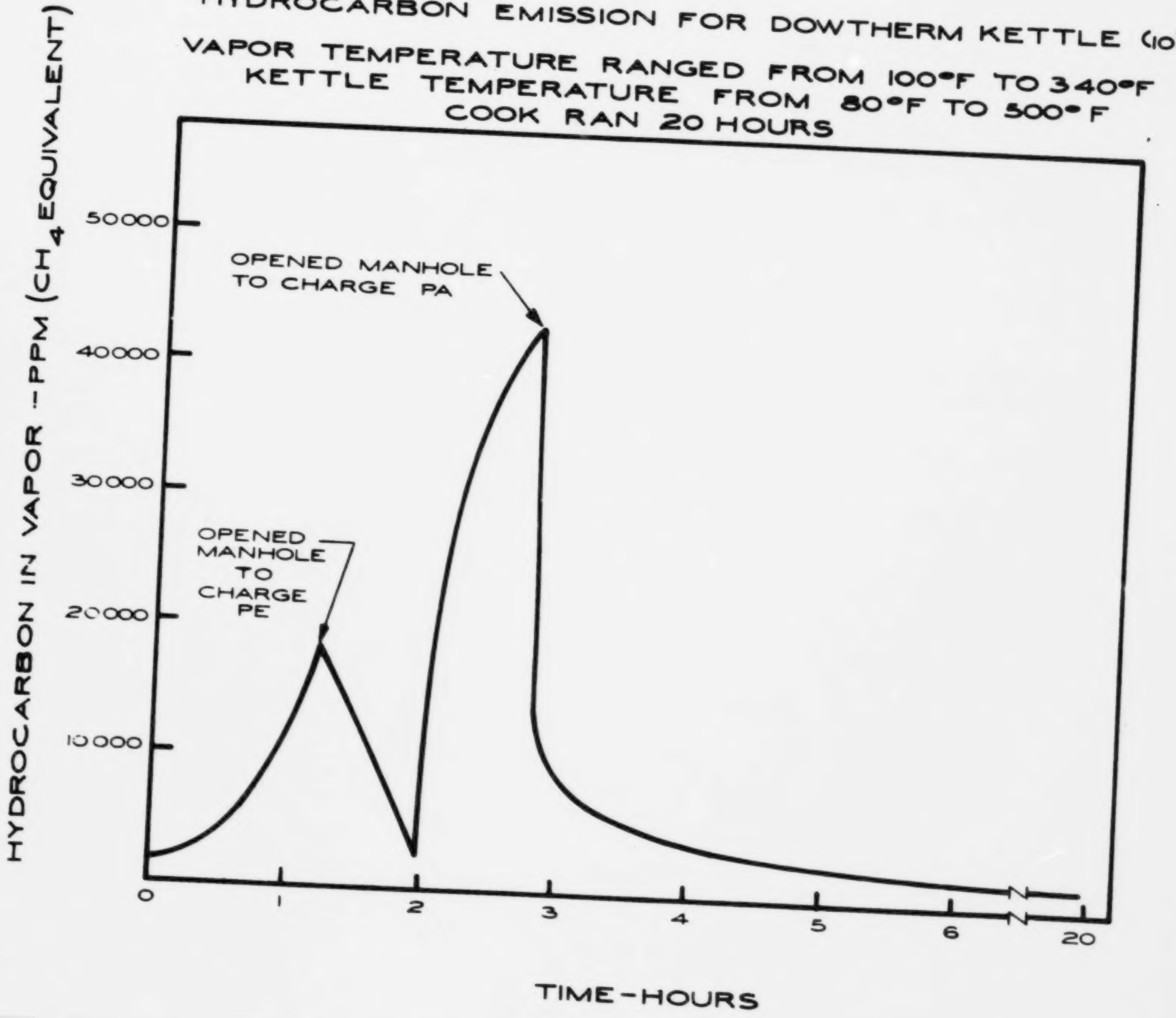


FIGURE 71

HYDROCARBON EMISSION FOR PAINT KETTLE (10)

VAPOR TEMPERATURE RANGED FROM 100° F TO 310° F
KETTLE TEMPERATURE FROM 80°F TO 480°F

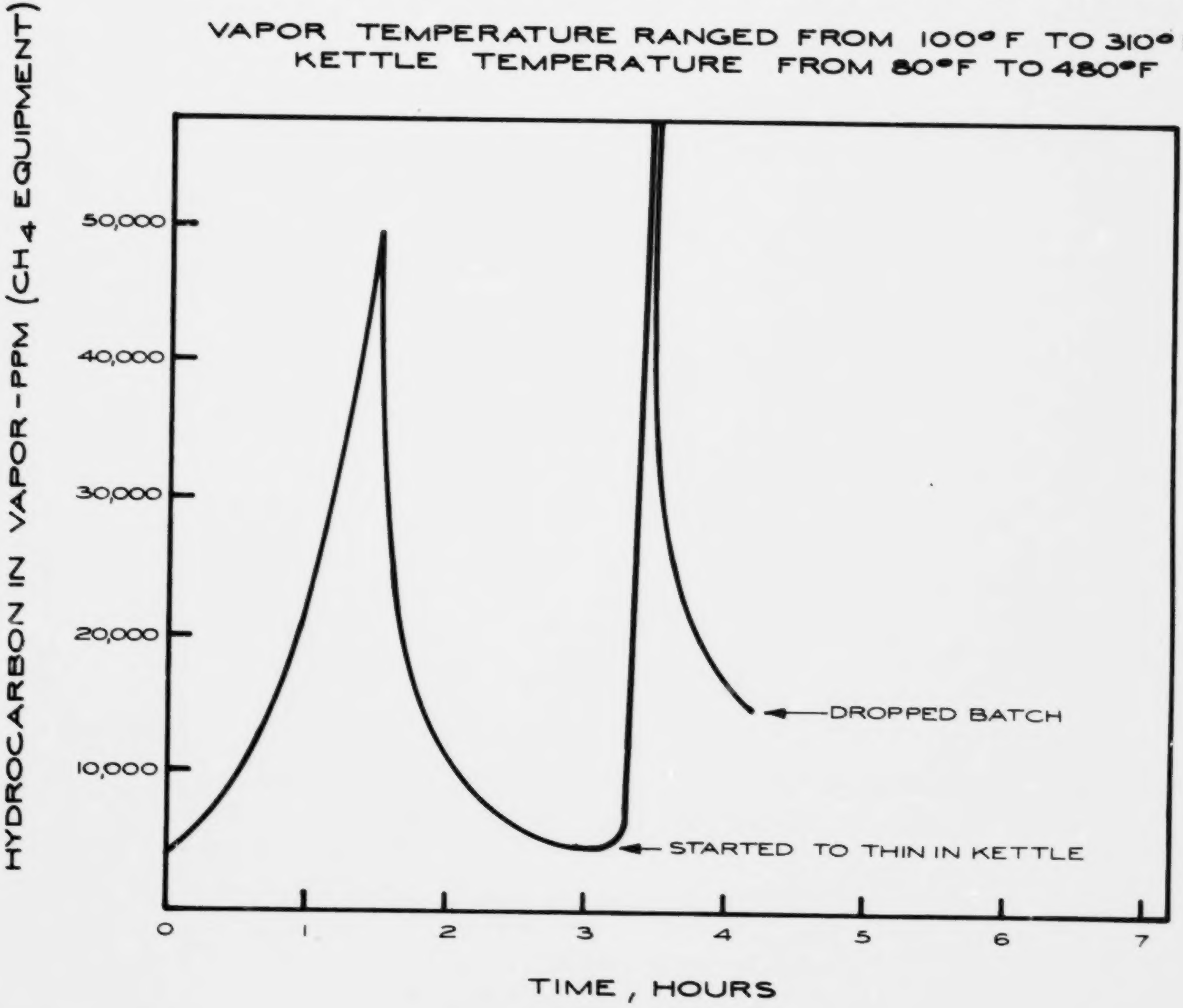
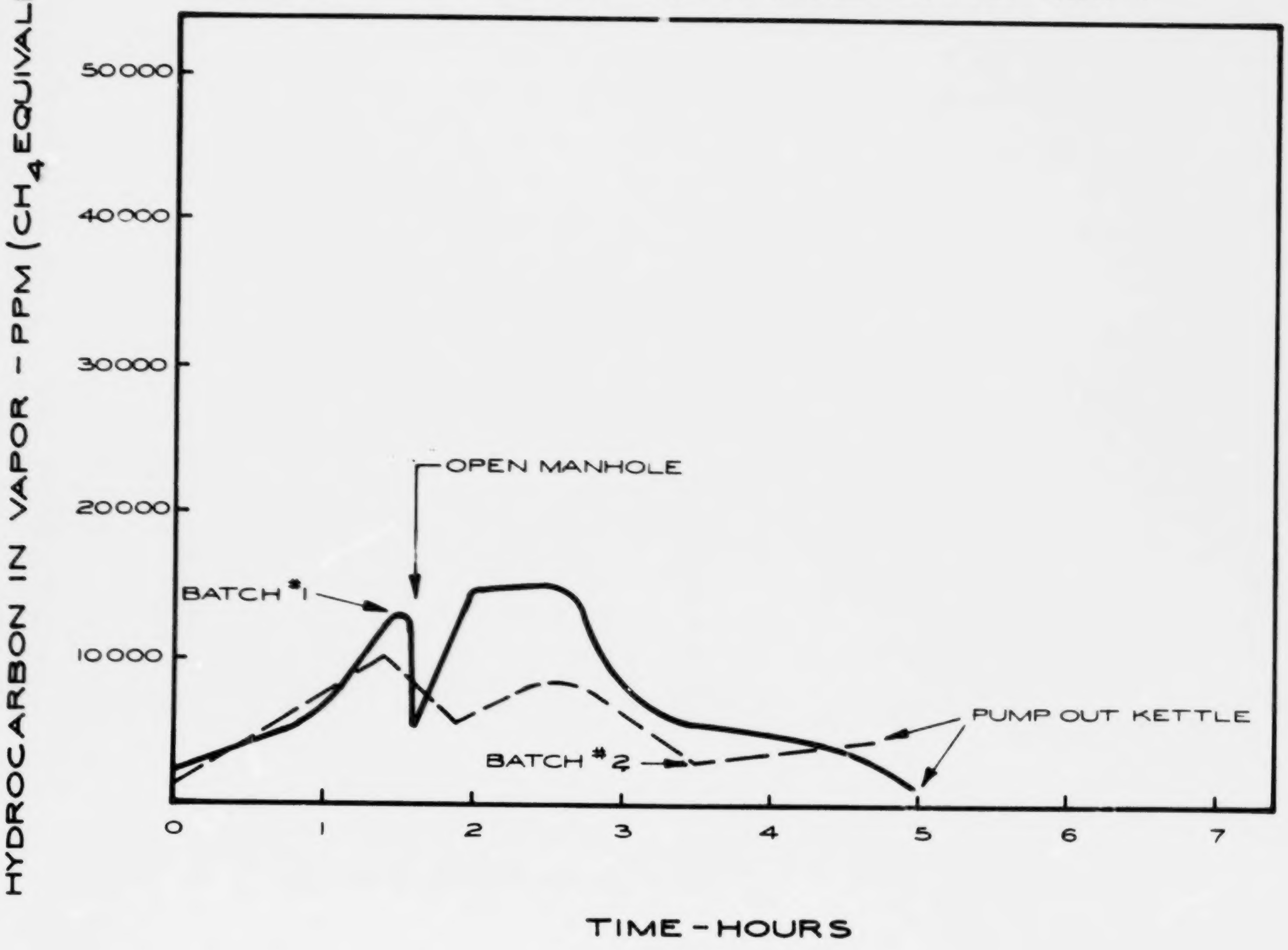


FIGURE 72

KETTLE HYDROCARBON EMISSON - 2 BATCHES MONITORED (10)

VAPOR TEMPERATURE RANGED FROM 100°F TO 310°F
KETTLE TEMPERATURE RANGED UP TO 490°F



The storage of liquid PA will result in significant vaporization losses from the storage tank and an effort must be made to control these losses. The most widely used method consists of an inert gas blanketing in conjunction with a pressure controlled unit. The tank is also equipped with a water cooled condenser used to vent the tank during filling. After filling the condenser is then heated with steam to remove collected PA by melting.

POLLUTION CONTROL CONSIDERATIONS

Collection of particulate pigment or resin emission is a simple straightforward job. The only practical control device is a fabric filter, and it is ideally suited for this application. Collection efficiency for the submicron pigment dust (0.05 to 0.25 microns) is in the range of 99.9%. There are no temperature problems since the exhaust system runs at ambient temperatures. The grain loading is very low and baglife is extensive. Approximately 0.01% of the loaded pigments are lost and collected. Grain loadings to the fabric filter run around 0.19 grain/SCF. A typical collection system is shown in Figure 73. The collection system can be a fixed hood which can handle both dust and pigment bags or a flexible hose positioned above the loading hatch or attached to the top of the tank. The tank attachment provides the most positive control of fugitive dust emission but also increases pigment and solvent losses slightly.

The application of control equipment to this problem is quite simple and can be solved with standard off-the-shelf equipment from a host of suppliers. For this reason, detailed equipment cost and installation bids will not be required.

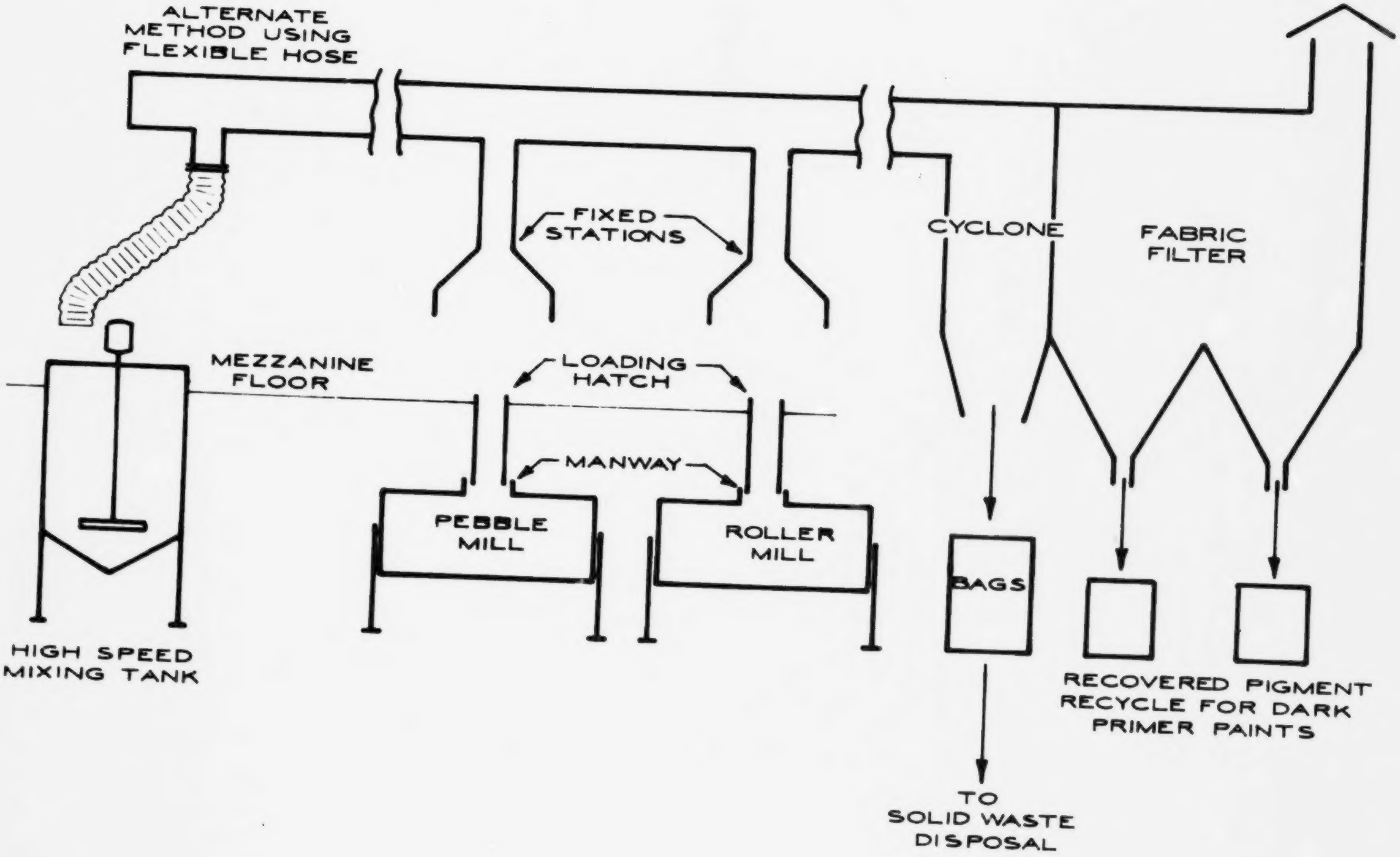
The control of hydrocarbon and odors from the various emission sources listed earlier is not quite as straightforward as the dust emission. There are three types of control equipment that have been applied to this problem. They are catalytic and thermal combustion devices, and wet scrubbers.

As a general rule, wet scrubbing does not provide a satisfactory solution for the following reasons:

1. Removal efficiency of fine hydrocarbon aerosol is not good at economically practical pressure drops.
2. Non-condensible hydrocarbon solvent vapors will not be removed.
3. Odor removal without the addition of an oxidizing agent such as

FIGURE 73

PIGMENT EMISSION CONTROL SYSTEM


ALTERNATE METHOD USING FLEXIBLE HOSE
FIXED STATIONS
CYCLONE
FABRIC FILTER
MEZZANINE FLOOR
LOADING HATCH
MANWAY
PEBBLE MILL
ROLLER MILL
BAGS
HIGH SPEED MIXING TANK
RECOVERED PIGMENT RECYCLE FOR DARK PRIMER PAINTS
TO SOLID WASTE DISPOSAL

potassium permanganate or sodium hypochorite is unsatisfactory. If an oxidizing agent is used, operating cost will be quite high due to the high concentration of other oxidizable material such as phthalic anydride, resins and oil.

4. Mobile packing and high make-up water rates are required to prevent plugging of the scrubber beds and spray nozzles.

5. Correction of the air pollution problem with wet scrubbing causes an equivalent water pollution problem which in many areas is more costly to correct than the original air pollution problem.

The only control technique currently being used that has proven effective for all cases is combustion. Three general methods are employed to combust waste gases:

1. Flame Combustion
2. Thermal Combustion
3. Catalytic Combustion

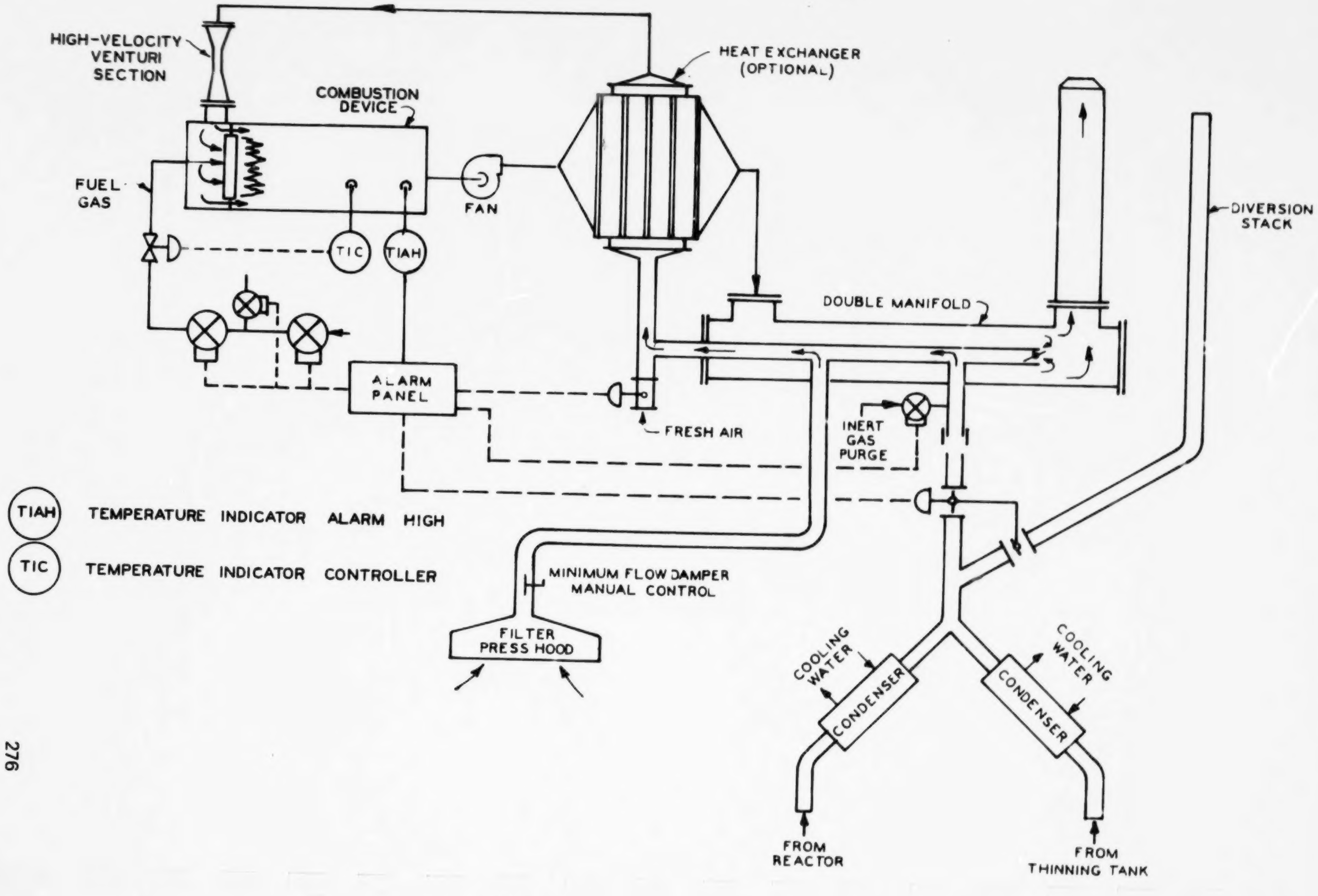
All of the above methods are oxidation processes. Ordinarily, each requires that the gaseous effluents be heated to the point where oxidation of the combustible will take place. The three methods differ basically in the temperature to which the gas stream must be heated.

Flame combustion is the easiest of the three to understand, as it comes the closest to everyday experience. When a gas stream is contaminated with combustibles at a concentration approaching the lower flammable limit, it is frequently practical to add a small amount of natural gas as an auxiliary fuel and sufficient air for combustion when necessary, and then pass the resulting mixture through a burner. The contaminants in the mixture serve as a part of the fuel. Flame incinerators of this type are most often used for closed chemical reactors. They are not used on resin reactors at present. They may be an ideal solution some day, however, when methods of operating a closed, pressurized resin reactor are developed.

It is far more likely that the concentration of combustible contaminants in an air stream will be well below the lower limit of flammability. When this is the case, direct thermal combustion is considerably more economical than flame combustion. Direct thermal combustion is carried out by equipment such as that illustrated in Figure 74. In this equipment, a gas burner is used to raise the temperature of the flowing stream sufficiently to cause a slow thermal reaction to occur in a residence chamber.

FIGURE 74

THERMAL COMBUSTION SYSTEM
FOR
RESIN REACTOR OR CLOSED KETTLE


HIGH-VELOCITY VENTURI SECTION
HEAT EXCHANGER (OPTIONAL)
COMBUSTION DEVICE
FUEL GAS
FAN
TIC
TIAH
DIVERSION STACK
DOUBLE MANIFOLD
ALARM PANEL
FRESH AIR
INERT GAS PURGE
TIAH TEMPERATURE INDICATOR ALARM HIGH
TIC TEMPERATURE INDICATOR CONTROLLER
MINIMUM FLOW DAMPER MANUAL CONTROL
FILTER PRESS HOOD
COOLING WATER
CONDENSER
COOLING WATER
CONDENSER
FROM REACTOR
FROM THINNING TANK

Whereas flame temperatures bring about oxidation by free radical mechanisms at temperatures of 2500° F and higher, thermal combustion of ordinary hydrocarbon compounds begins to take place at temperatures as low as 900 to 1000° F. Good conversion efficiencies are produced at temperatures in the order of 1400° F with a residence time of 0.3 to 0.6 seconds.

Catalytic combustion is carried out by bringing the gas stream into intimate contact with a bed of catalyst. In this sytem, the reaction takes place directly upon the surface of the catalyst, which is usually composed of precious metals, such as platinum and palladium. While thermal combustion equipment brings about oxidation at concentrations below the limits of flame combustion, catalytic combustion takes place below the limits of flammability and below the normal oxidation temperatures of the contaminants. The reaction is instantaneous by comparison to thermal combustion and no residence chamber is required. Catalytic combustion is carried out by equipment such as that illustrated in Figure 75.

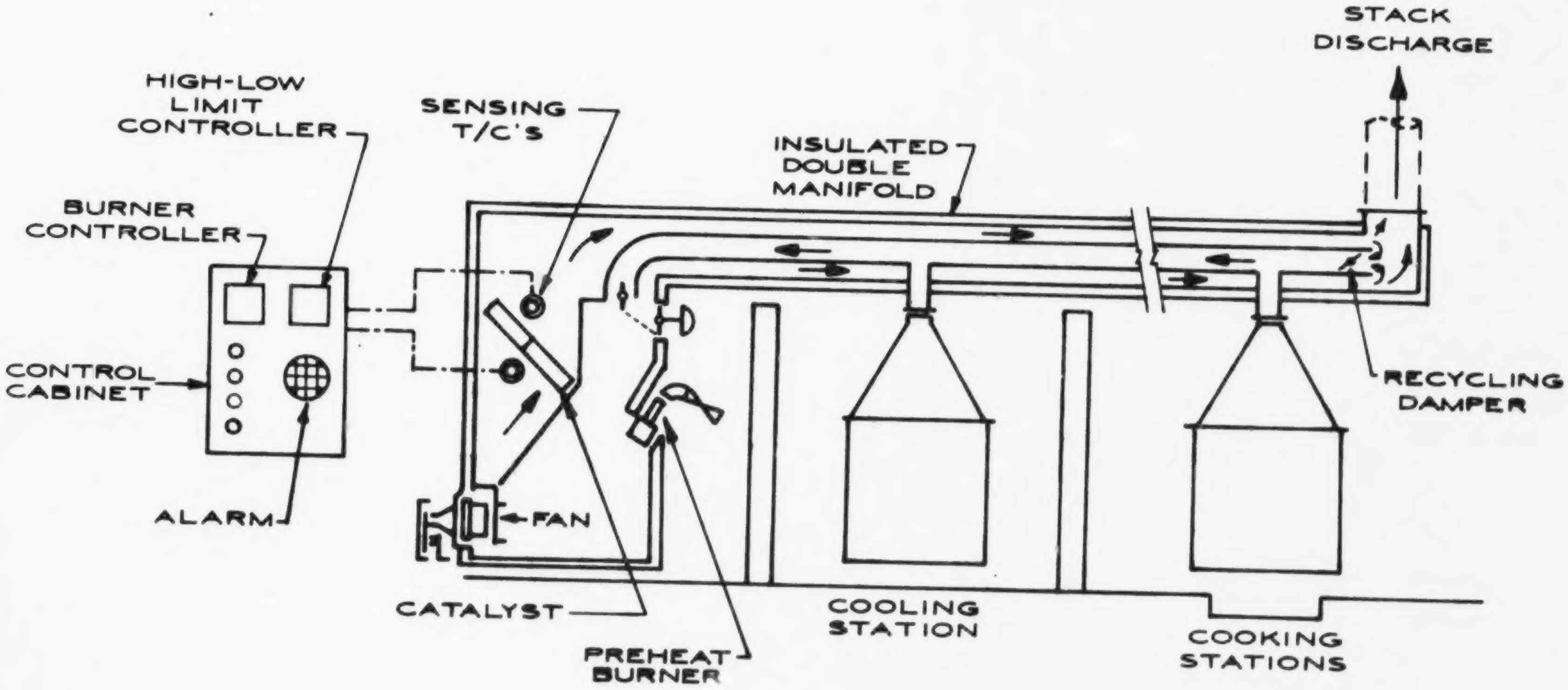
In general, catalytic afterburners are less expensive to operate. However, they depend directly on the performance of the catalyst for their effectiveness. They will not function properly if the catalyst becomes deactivated. Because of this, catalytic units are not inherently functional when operated at design conditions. In many areas, means for ensuring adequate performance of the catalyst on a long term basis will be required by environmental control offices.

The basis for design of either catalytic or thermal combustion is the hydrocarbon concentration of the exhaust gases handled by the incinerator. The maximum hydrocarbon level is set by most insurance companies at one-quarter of the lower explosive limit (LEL) which is equivalent to 13 Btu/SCF of exhaust gas. As outlined earlier, the quantity of emission may vary significantly with cooking time and the type of cook. There is also likely to be a very large variety of different hydrocarbons emitted. For this reason, theoretical calculations of emission for design purposes are not satisfactory. On site emission measurements, as shown earlier on Figures 70 and 71, are required. Once the rate of emission is determined, it is then necessary to calculate the dilution air required to meet 1/4 LEL and set up the duct work system to provide for this dilution. When possible, dilution air should be utilized to help capture as many fugitive fume emissions as possible. For example this can be accomplished by taking the dilution air from a hood positioned over the resin filter press and venting the thinning tanks and product run down tanks into the same system.

A concentration of 1/4 LEL or 13 Btu/SCF will give a temperature rise of

FIGURE 75

SCHEMATIC DIAGRAM OF A CATALYTIC COMBUSTION SYSTEM FOR VARNISH KETTLES


STACK DISCHARGE
HIGH-LOW LIMIT CONTROLLER
SENSING T/C'S
INSULATED DOUBLE MANIFOLD
BURNER CONTROLLER
CONTROL CABINET
RECYCLING DAMPER
ALARM
FAN
CATALYST
COOLING STATION
COOKING STATIONS
PREHEAT BURNER

about 600°F in the afterburner. This is too high if a heat exchanger is to be used, and in these cases, dilution will be required to a maximum concentration of 12 Btu/SCF. In all cases, the heat exchangers will be the parallel flow type having a thermal efficiency of 42%. This is required to assure temperature balance and control, due to the high emission concentration.

The major problem with catalytic or thermal afterburners as applied to open or closed resin and varnish kettles is the danger of fires and/or explosions. This has happened in numerous occasions in the past due primarily to excessive hydrocarbon emission from kettles. These problems have been all but eliminated on newer units by assuring that the design was based on actual emission measurements of the highest emitting cook and the addition of some of the following system safety features:

1. High limit temperature alarm to shut off burner and activate a diversion system.
2. High velocity duct section to assure gas flow to afterburner substantially exceeds flame propogation velocity of hydrocarbons being burned.
3. Double manifolding or hot gas recycle to prevent condensation of heavy hydrocarbons or phthalic anhydride.
4. Diversion system to block off hydrocarbon emissions to unit, by-passing them directly out of separate exhaust, and introduce fresh air to purge the unit.
5. Pneumatic operation of the diversion system to assure fast positive action and provide a fail safe system in the event of either air or electrical failure.
6. Purging with inert gas in the event of power failure.

The above general requirements are applicable to all types of afterburner control. Specific details for each type of system will be given in the equipment specifications.

SPECIFICATIONS AND COSTS

Specifications were written for combustion systems applied to two sources typical of those in the paint and varnish industry: closed resin reactors and open cooking kettles. Each of these sources emits primarily hydrocarbons. Both thermal and catalytic combustion systems were specified for abatement in each application. For those applications where the gas flow was sufficient, separate, but parallel, specifications were written for combustion systems with and without heat exchange. Separate specifications and cost tables are presented for each application.

For the graphical presentation of cost data, however, sets of similar data were grouped together rather than showing one set of cost graphs for each application. This grouping permits the presentation of cost graphs covering a wider range of combustion system sizes. Operating cost curves for similar thermal and catalytic systems are presented on the same figure for comparison purposes. The data were grouped and presented as follows:

System Type	Capital Cost Figure	Operating Cost Figure
Thermal, without heat exchange	26	28, 29
Catalytic, without heat exchange	27	28, 29
Thermal, with heat exchange	30	32, 33
Catalytic, with heat exchange	31	32, 33

In addition to the cost data gathered under this contract, similar cost data gathered at the same time under contract No. 68-02-0259, Air Pollution Control Engineering and Cost Study of the Paint and Varnish Industry, are also presented on these figures. These additional data are not included in the average cost tables.

Preceding page blank

TABLE 93

THERMAL INCINERATOR PROCESS DESCRIPTION FOR RESIN REACTOR SPECIFICATION

This specification describes the requirements for a thermal combustion system for abatement of the hydrocarbon emissions from the resin production facility of a paint and varnish plant. The system will be similar to that shown in Figure 24. All reactor thinning tanks and product rundown tanks will be vented to the collection system. Dilution air will be supplied through a hood over the resin filter press(es). A minimum flow damper is to be supplied in this part of the ventilation system. It is to be sized to allow for a maximum fume concentration of 40% LEL in the fully closed position. A high velocity Venturi section is to be located at the incinerator inlet to assure gas flow is in excess of flame propagation velocity at 1/2 design flow rate.

The afterburner is to be natural gas fired. Sufficient gas, having a specific gravity of 0.60 and an upper heating value of 1040 Btu/SCF, is available at pressure of 1.0 psig.

The exhaust gas contains sufficient oxygen, greater than 16% O_2*, to allow firing of the afterburner with a raw gas or process air burner. A combustion air system is not required. Fume load to the incinerator is composed of 75% kettle emission and 25% from other sources. Average emissions are 60% of peak. The system is to be designed on peak emission but operating costs are to be based on average emissions. Operating conditions listed are based on peak emissions. Fume destruction in burner is to be calculated as follows:*

10% for catalytic units with or without heat exchange

10% for thermal units with heat exchange

20% for thermal units without heat exchange.

Please fill in estimated efficiency of afterburner and burner duty.

The afterburner is to be supplied with a suitable control panel and all equipment is to be designed for outdoor operation. Incinerator operating and safety controls are to be designed to meet F.I.A. (Factory Insurance Association) requirements. All dampers are to be pneumatically operated and contain an integral fail-safe air reservoir. The system is to automatically divert in the event of low flow, high afterburner temperature, high reactor pressure, and afterburner preheat burner failure. The exhaust system should also be purged with inert gas on fan failure or loss of flow. Damper operating is to be sequential with the position switches mounted on the damper arm and not the operator. The system fan shall be located after the preheat burner or the incinerator outlet and shall be constructed to withstand 200°F higher than design operating temperature. The fan shall have a V-belt drive and fan and motor shall have the capacity to overcome the pressure drop of the ductwork, afterburner and any heat exchanger that may be used. System ductwork should be sized for a maximum ΔP *of 2 in. w.c. hot. The fan motor may be sized for restricted flow cold start. The ductwork to the incinerator should be heated either by the use of a double manifold or hot gas recycle or a combination of both.*

INSTALLATION

A complete turnkey proposal including ductwork, structural steel, fuel and inert gas piping, etc., is requested. For the purpose of this proposal fan and damper including operators are to be considered as auxiliary. The controls and control cabinet are to be included with the afterburner price. The afterburner will be assumed to be located on a structural steel base on the resin plant roof. No modification to the building structural steel is required. A tie through the roof from the base to the building steel is required. The base and tie-in are part of the installed structural cost. All utilities are available within 30 ft of the control cabinet, motor, and burner. Plant air is available at 100 psig and located within 30 ft. Inert gas is available and will require 30 ft of piping from supply to ductwork. The existing stack can be used for the diversion stack. A 10 ft exhaust stack will be mounted on the incinerator, double manifold or exhaust fan. A total of thirty-five (35) feet of new exhaust ductwork and twenty-five (25) feet of double manifold will be required.

TABLE 94

THERMAL INCINERATOR OPERATING CONDITIONS FOR RESIN REACTOR SPECIFICATION (WITHOUT HEAT EXCHANGE)

	Small	*Large*
Process Conditions		
Reactors, Number	*1*	*3*
Size each, gal	*4,000*	*5,000*
Hcbn Emission Max, lb/hr	*95*	*317*
Hcbn Emission Ave, lb/hr	*57*	*190*
Total Exhaust Rate, SCFM	*3,000*	*10,000*
Exhaust Temperature, ^oF	*110*	*110*
Heat of Combustion of Reactor Fume, Btu/lb	*17,000*	*17,000*
Hydrocarbon Concentration		
Maximum, Btu/SCF	*12*	*12*
Average, Btu/SCF	*7*	*7*
Incinerator Without Heat Exchange		
Unit Inlet	*325*	*325*
Burner ΔT from Fuel Gas, ^oF	*550*	*550*
**Burner ΔT from Flame Combustion, ^oF*	*125*	*125*
Burner Outlet Temperature, ^oF	*1,000*	*1,000*
***Unit ΔT from Thermal Combustion, ^oF*	*435*	*435*
Unit Outlet Temperature, ^oF	*1,435*	*1,435*
****Burner Duty, MM Btu/hr*	*3.36*	*11.19*

* *Assumes 20% fume combustion in burner flame*

** *Assumes 95% overall fume combustion*

*** *Supplied by bidders*

Preceding page blank

TABLE 95

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR RESIN REACTORS (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			3,225	10,750
°F			110	110
SCFM			3,000	10,000
Moisture Content, Vol. %			4.3	4.3
Effluent Contaminant Loading				
lb/hr, Max			95	317
lb/hr, Avg			57	190
Cleaned Gas Flow				
ACFM			10,840	36,130
°F			1,435	1,435
SCFM			3,036	10,120
Moisture Content, Vol. %			9.2	9.2
Cleaned Gas Contaminant Loading				
lb/hr			4.8	16
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			14,806	27,071
(2) Auxiliaries Cost				
(a) Fan(s)			1,636	4,655
(b) Pump(s)				
(c) Damper(s)			1,085	1,405
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			1,800	2,650
(b) Foundations & Support			1,913	2,725
Ductwork			2,600	4,150
Stack			850	1,750
Electrical			600	1,100
Piping			510	720
Insulation				
Painting				
Supervision			900	900
Startup			400	400
Performance Test			1,500	1,500
Other				
(4) Total Cost			28,600	49,026

TABLE 96

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR THERMAL INCINERATORS FOR RESIN REACTORS (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,760				
Operating Labor (if any)					
Operator	$6/hr			375	375
Supervisor				-	-
Total Operating Labor				375	375
Maintenance					
Labor	$6/hr			390	390
Materials				160	210
Total Maintenance				550	600
Replacement Parts					
Total Replacement Parts				375	475
Utilities					
Electric Power	$.011/kw-hr			2,024	5,059
Fuel	$.80/MM Btu			20,784	69,026
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				22,808	74,085
Total Direct Cost				24,108	75,535
Annualized Capital Charges				2,860	4,903
Total Annual Cost				26,968	80,438

TABLE 97

THERMAL INCINERATOR PROCESS DESCRIPTION FOR OPEN KETTLE SPECIFICATION

This specification describes the requirements for a thermal combustion system for abatement of the hydrocarbon emissions from the resin production facility of a paint and varnish plant. The system will be similar to that shown in Figure 24. All reactor thinning tanks and product rundown tanks will be vented to the collection system. Dilution air will be supplied through a hood over the resin filter press(es). A minimum flow damper is to be supplied in this part of the ventilation system. It is to be sized to allow for a maximum fume concentration of 40% LEL in the fully closed position. A high velocity Venturi section is to be located at the incinerator inlet to assure gas flow is in excess of flame propagation velocity at 1/2 design flow rate.

The afterburner is to be natural gas fired. Sufficient gas, having a specific gravity of 0.60 and an upper heating value of 1040 Btu/SCF, is available at pressure of 1.0 psig.

The exhaust gas contains sufficient oxygen, greater than 16% O_2*, to allow firing of the afterburner with a raw gas or process air burner. A combustion air system is not required. Fume load to the incinerator is composed of 75% kettle emission and 25% from other sources. Average emissions are 60% of peak. The system is to be designed on peak emission but operating costs are to be based on average emissions. Operating conditions listed are based on peak emissions. Fume destruction in burner is to be calculated as follows:*

10% for catalytic units with or without heat exchange

10% for thermal units with heat exchange

20% for thermal units without heat exchange.

Please fill in estimated efficiency of afterburner and burner duty.

The afterburner is to be supplied with a suitable control panel and all equipment is to be designed for outdoor operation. Incinerator operating and safety controls are to be designed to meet F.I.A. (Factory Insurance Association) requirements. All dampers are to be pneumatically operated and contain an integral fail-safe air reservoir. The system is to automatically divert in the event of low flow, high afterburner temperature, high reactor pressure, and afterburner preheat burner failure. The exhaust system should also be purged with inert gas on fan failure or loss of flow. Damper operating is to be sequential with the position switches mounted on the damper arm and not the operator. The system fan shall be located after the preheat burner or the incinerator outlet and shall be constructed to withstand 200°F higher than design operating temperature. The fan shall have a V-belt drive and fan and motor shall have the capacity to overcome the pressure drop of the ductwork, afterburner and any heat exchanger that may be used. System ductwork should be sized for a maximum ΔP *of 2 in. w.c. hot. The fan motor may be sized for restricted flow cold start. The ductwork to the incinerator should be heated either by the use of a double manifold or hot gas recycle or a combination of both.*

INSTALLATION

A complete turnkey proposal including ductwork, structural steel, fuel and inert gas piping, etc., is requested. For the purpose of this proposal fan and damper including operators are to be considered as auxiliary. The controls and control cabinet are to be included with the afterburner price. The afterburner will be assumed to be located on a structural steel base on the resin plant roof. No modification to the building structural steel is required. A tie through the roof from the base to the building steel is required. The base and tie-in are part of the installed structural cost. All utilities are available within 30 ft of the control cabinet, motor, and burner. Plant air is available at 100 psig and located within 30 ft. Inert gas is available and will require 30 ft of piping from supply to ductwork. The existing stack can be used for the diversion stack. A 10 ft exhaust stack will be mounted on the incinerator, double manifold or exhaust fan. A total of thirty-five (35) feet of new exhaust ductwork and twenty-five (25) feet of double manifold will be required.

TABLE 98

THERMAL INCINERATOR OPERATING CONDITIONS FOR OPEN KETTLE SPECIFICATION (WITHOUT HEAT EXCHANGE)

	Small	Large
Process Conditions		
Kettles, Number	1	3
Size each, gal	200	375
Hcbn Emission Max, lb/hr	16.8	101.1
Hcbn Emission Ave, lb/hr	10.1	60.6
Total Exhaust Rate, SCFM	500	3,000
Exhaust Temperature, °F	80	80
Heat of Combustion of Kettle Fume, Btu/lb	16,000	16,000
Hydrocarbon Concentration Maximum, Btu/SCF	12	12
Average, Btu/SCF	7	7
Incinerator Without Heat Exchange		
Unit Inlet	300	300
Burner ΔT from Fuel Gas, °F	575	575
*Burner ΔT from Flame Combustion, °F	125	125
Burner Outlet Temperature, °F	1,000	1,000
**Unit ΔT from Thermal Combustion, °F	435	435
Unit Outlet Temperature, °F	1,435	1,435
***Burner Duty, MM Btu/hr	0.58	3.43

* Assumes 20% fume combustion in burner flame

** Assumes 95% overall fume combustion

*** Supplied by bidders

Preceding page blank

TABLE 99

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR OPEN KETTLES (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			509	3,056
°F			80	80
SCFM			500	3,000
Moisture Content, Vol. %			1.7	1.7
Effluent Contaminant Loading				
lb/hr, Max			16.8	101.1
lb/hr, Avg			10.1	60.6
Cleaned Gas Flow				
ACFM			1,810	10,860
°F			1,435	1,435
SCFM			506	3,036
Moisture Content, Vol. %			6.7	6.7
Cleaned Gas Contaminant Loading				
lb/hr			8.4	50.5
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			12,081	14,806
(2) Auxiliaries Cost				
(a) Fan(s)			720	1,636
(b) Pump(s)				
(c) Damper(s)			923	1,085
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			1,650	1,913
(b) Foundations & Support			1,650	1,913
Ductwork			2,360	3,025
Stack			750	850
Electrical			550	600
Piping			470	510
Insulation				
Painting				
Supervision			800	900
Startup			350	350
Performance Test			1,150	1,250
Other				
(4) Total Cost			23,454	28,838

TABLE 100

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR THERMAL INCINERATORS FOR OPEN KETTLES (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,760				
Operating Labor (if any)					
Operator	$6/hr			375	375
Supervisor				-	-
Total Operating Labor				375	375
Maintenance					
Labor	$6/hr			390	390
Materials				110	160
Total Maintenance				500	550
Replacement Parts					
Total Replacement Parts				325	375
Utilities					
Electric Power	$.011/kw-hr			482	2,024
Fuel	$.80/MM Btu			3,554	21,003
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				4,036	23,027
Total Direct Cost				5,236	24,327
Annualized Capital Charges				2,345	2,884
Total Annual Cost				7,581	27,211

TABLE 101

CATALYTIC INCINERATOR PROCESS DSCRIPTION FOR RESIN REACTOR SPECIFICATION

This specification describes the requirements for a catalytic combustion system for abatement of the hydrocarbon emissions from the resin production facility of a paint and varnish plant. The system will be similar to that shown in Figure 24. All reactor thinning tanks and product rundown tanks will be vented to the collection system. Dilution air will be supplied through a hood over the resin filter press(es). A minimum flow damper is to be supplied in this part of the ventilation system. It is to be sized to allow for a maximum fume concentration of 40% LEL in the fully closed position. A high velocity Venturi section is to be located at the incinerator inlet to assure gas flow is in excess of flame propagation velocity at 1/2 design flow rate.

The afterburner is to be natural gas fired. Sufficient gas, having a specific gravity of 0.60 and an upper heating value of 1040 Btu/SCF, is available at pressure of 1.0 psig.

The exhaust gas contains sufficient oxygen, greater than 16% 0_2, *to allow firing of the afterburner with a raw gas or process air burner. A combustion air system is not required. Fume load to the incinerator is composed of 75% kettle emission and 25% from other sources. Average emissions are 60% of peak. The system is to be designed on peak emission but operating costs are to be based on average emissions. Operating conditions listed are based on peak emissions. Fume destruction in burner is to be calculated as follows:*

10% for catalytic units with or without heat exchange

10% for thermal units with heat exchange

20% for thermal units without heat exchange.

Please fill in estimated efficiency of afterburner, burner duty, catalyst face velocity, and catalyst volume.

The afterburner is to be supplied with a suitable control panel and all equipment is to be designed for outdoor operation. Incinerator operating and safety controls are to be designed to meet F.I.A. (Factory Insurance Association) requirements. All dampers are to be pneumatically operated and contain an integral fail-safe air reservoir. The system is to automatically divert in the event of low flow, high afterburner temperature, high reactor pressure, and afterburner preheat burner failure. The exhaust system should also be purged with inert gas on fan failure or loss of flow. Damper operating is to be sequential with the position switches mounted on the damper arm and not the operator. The system fan shall be located after the preheat burner or the incinerator outlet and shall be constructed to withstand 200°F higher than design operating temperature. The fan shall have a V-belt drive and fan and motor shall have the capacity to overcome the pressure drop of the ductwork, afterburner and any heat exchanger that may be used. System ductwork should be sized for a maximum Δ *P of 2 in. w.c. hot. The fan motor may be sized for restricted flow cold start. The ductwork to the incinerator should be heated either by the use of a double manifold or a combination of both.*

INSTALLATION

A complete turnkey proposal including ductwork, structural steel, fuel and inert gas piping, etc., is requested. For the purpose of this proposal fan and damper including operators are to be considered as auxiliary. The controls and control cabinet are to be included with the afterburner price. The afterburner will be assumed to be located on a structural steel base on the resin plant roof. No modification to the building structural steel is required. A tie through the roof from the base to the building steel is required. The base and tie-in are part of the installed structural cost. All utilities are available within 30 ft of the control cabinet, motor, and burner. Plant air is available at 100 psig and located within 30 ft. Inert gas is available and will require 30 ft of piping from supply to ductwork. The existing stack can be used for the diversion stack. A 10 ft exhaust stack will be mounted on the incinerator, double manifold or exhaust fan. A total of thirty-five (35) feet of new exhaust ductwork and twenty-five (25) feet of double manifold will be required.

TABLE 102

CATALYTIC INCINERATOR OPERATING CONDITIONS FOR RESIN REACTOR SPECIFICATION (WITHOUT HEAT EXCHANGE)

	Small	*Large*
Process Conditions		
Reactors, Number	*1*	*3*
Size each, gal	*4,000*	*5,000*
Hcbn Emission Max, lb/hr	*95*	*317*
Hcbn Emission Ave, lb/hr	*57*	*190*
Total Exhaust Rate, SCFM	*3,000*	*10,000*
Exhaust Temperature, °F	*110*	*110*
Heat of Combustion of Reactor Fume, Btu/lb	*17,000*	*17,000*
Hydrocarbon Concentration Maximum, Btu/SCF	*12*	*12*
Average, Btu/SCF	*7*	*7*
Incinerator Without Heat Exchange		
Unit Inlet	*300*	*300*
Burner ΔT from Fuel Gas, °F	*240*	*240*
**Burner ΔT from Flame Combustion, °F*	*60*	*60*
Burner Outlet Temperature, °F	*600*	*600*
***Unit ΔT from Thermal Combustion, °F*	*500*	*500*
Unit Outlet Temperature, °F	*1,100*	*1,100*
Burner Duty, MM Btu/hr	*– – –*	*– – –*

* *Assumes 10% fume combustion in burner flame*

** *Assumes 95% overall fume combustion*

Preceding page blank

TABLE 103

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CATALYTIC INCINERATORS FOR RESIN REACTORS (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			3,225	10,750
°F			110	110
SCFM			3,000	10,000
Moisture Content, Vol. %			4.3	4.3
Effluent Contaminant Loading				
lb/hr, Max			95	317
lb/hr, Avg			57	190
Cleaned Gas Flow				
ACFM			8,880	29,600
°F			1,100	1,100
SCFM			3,015	1,000
Cleaned Gas Contaminant Loading				
lb/hr			4.8	15.9
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			22,988	43,260
(2) Auxiliaries Cost (a) Fan(s) (b) Pump(s) (c) Damper(s) (d) Conditioning, Equipment (e) Dust Disposal Equipment			3,990	8,515
(3) Installation Cost (a) Engineering (b) Foundations & Support Ductwork Stack Electrical Piping Insulation Painting Supervision Startup Performance Test Other			11,386	18,123
(4) Total Cost			38,364	69,898

TABLE 104

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR CATALYTIC INCINERATORS FOR RESIN REACTORS (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,760				
Operating Labor (if any)					
Operator	$6/hr			675	675
Supervisor	$8/hr			200	200
Total Operating Labor				875	875
Maintenance					
Labor	$6/hr			510	510
Materials				115	145
Total Maintenance				625	655
Replacement Parts					
Total Replacement Parts				1,265	4,045
Utilities					
Electric Power	$.011/kw-hr			1,118	3,239
Fuel	$.80/MMBtu			9,251	30,835
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				10,369	34,074
Total Direct Cost				13,134	39,649
Annualized Capital Charges				3,836	6,990
Total Annual Cost				16,970	46,639

TABLE 105

CATALYTIC INCINERATOR PROCESS DESCRIPTION FOR OPEN KETTLE SPECIFICATION

This specification describes the requirements for a catalytic combustion system for abatement of the hydrocarbon emissions from the resin production facility of a paint and varnish plant. The system will be similar to that shown in Figure 24. All reactor thinning tanks and product rundown tanks will be vented to the collection system. Dilution air will be supplied through a hood over the resin filter press(es). A minimum flow damper is to be supplied in this part of the ventilation system. It is to be sized to allow for a maximum fume concentration of 40% LEL in the fully closed position. A high velocity Venturi section is to he located at the incinerator inlet to assure gas flow is in excess of flame propagation velocity at 1/2 design flow rate.

The afterburner is to be natural gas fired. Sufficient gas, having a specific gravity of 0.60 and an upper heating value of 1040 Btu/SCF, is available at pressure of 1.0 psig.

The exhaust gas contains sufficient oxygen, greater than 16% O_2, *to allow firing of the afterburner with a raw gas or process air burner. A combustion air system is not required. Fume load to the incinerator is composed of 75% kettle emission and 25% from other sources. Average emissions are 60% of peak. The system is to be designed on peak emission but operating costs are to be based on average emissions. Operating conditions listed are based on peak emissions. Fume destruction in burner is to be calculated as follows:*

10% for catalytic units with or without heat exchange

10% for thermal units with heat exchange

20% for thermal units without heat exchange.

Please fill in estimated efficiency of afterburner, burner duty, catalyst face velocity, and catalyst volume.

The afterburner is to be supplied with a suitable control panel and all equipment is to be designed for outdoor operation. Incinerator operating and safety controls are to be designed to meet F.I.A. (Factory Insurance Association) requirements. All dampers are to be pneumatically operated and contain an integral fail-safe air reservoir. The system is to automatically divert in the event of low flow, high afterburner temperature, high reactor pressure, and afterburner preheat burner failure. The exhaust system should also be purged with inert gas on fan failure or loss of flow. Damper operating is to be sequential with the position switches mounted on the damper arm and not the operator. The system fan shall be located after the preheat burner or the incinerator outlet and shall be constructed to withstand 200°F higher than design operating temperature. The fan shall have a V-belt drive and fan and motor shall have the capacity to overcome the pressure drop of the ductwork, afterburner and any heat exchanger that may be used. System ductwork should be sized for a maximum ΔP *of 2 in. w.c. hot. The fan motor may be sized for restricted flow cold start. The ductwork to the incinerator should be heated either by the use of a double manifold or hot gas recycle or a combination of both.*

INSTALLATION

A complete turnkey proposal including ductwork, structural steel, fuel and inert gas piping, etc., is requested. For the purpose of this proposal fan and damper including operators are to be considered as auxiliary. The controls and control cabinet are to be included with the afterburner price. The afterburner will be assumed to be located on a structural steel base on the resin plant roof. No modification to the building structural steel is required. A tie through the roof from the base to the building steel is required. The base and tie-in are part of the installed structural cost. All utilities are available within 30 ft of the control cabinet, motor, and burner. Plant air is available at 100 psig and located within 30 ft. Inert gas is available and will require 30 ft of piping from supply to ductwork. The existing stack can be used for the diversion stack. A 10 ft exhaust stack will be mounted on the incinerator, double manifold or exhaust fan. A total of thirty-five (35) feet of new exhaust ductwork and twenty-five (25) feet of double manifold will be required.

TABLE 106

CATALYTIC INCINERATOR OPERATING CONDITIONS FOR OPEN KETTLE SPECIFICATION (WITHOUT HEAT EXCHANGE)

	Small	*Large*
Process Conditions		
Kettles Number	1	3
Size each, gal	200	375
Hcbn Emission Max, lb/hr	16.8	101.1
Hcbn Emission Ave, lb/hr	10.1	60.6
Total Exhaust Rate, SCFM	500	3,000
Exhaust Temperature, °F	80	80
Heat of Combustion of Kettle Fume, Btu/lb	16,000	16,000
Hydrocarbon Concentration		
Maximum, Btu/SCF	12	12
Average, Btu/SCF	7	7
Incinerator Without Heat Exchange		
Unit Inlet	280	280
Burner ΔT from Fuel Gas, °F	260	260
**Burner ΔT from Flame Combustion, °F*	60	60
Burner Outlet Temperature, °F	600	600
***Unit ΔT from Thermal Combustion, °F*	500	500
Unit Outlet Temperature, °F	1,100	1,100
Burner Duty, MM Btu/hr	– – –	– – –

* *Assumes 10% fume combustion in burner flame*
** *Assumes 95% overall fume combustion*

Preceding page blank

TABLE 107

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CATALYTIC INCINERATORS FOR OPEN KETTLES (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			509	3,056
°F			80	80
SCFM			500	3,000
Moisture Content, Vol. %			1.7	1.7
Effluent Contaminant Loading				
lb/hr, Max			16.8	101.1
lb/hr, Avg			10.1	60.6
Cleaned Gas Flow				
ACFM			1,480	8,880
°F			1,100	1,100
SCFM			503	3,018
Moisture Content, Vol. %			5.4	5.4
Cleaned Gas Contaminant Loading				
lb/hr			0.8	5.0
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			11,974	22,988
(2) Auxiliaries Cost (a) Fan(s) (b) Pump(s) (c) Damper(s) (d) Conditioning, Equipment (e) Dust Disposal Equipment			2,218	3,990
(3) Installation Cost (a) Engineering (b) Foundations & Support Ductwork Stack Electrical Piping Insulation Painting Supervision Startup Performance Test Other			9,003	11,773
(4) Total Cost			23,195	38,751

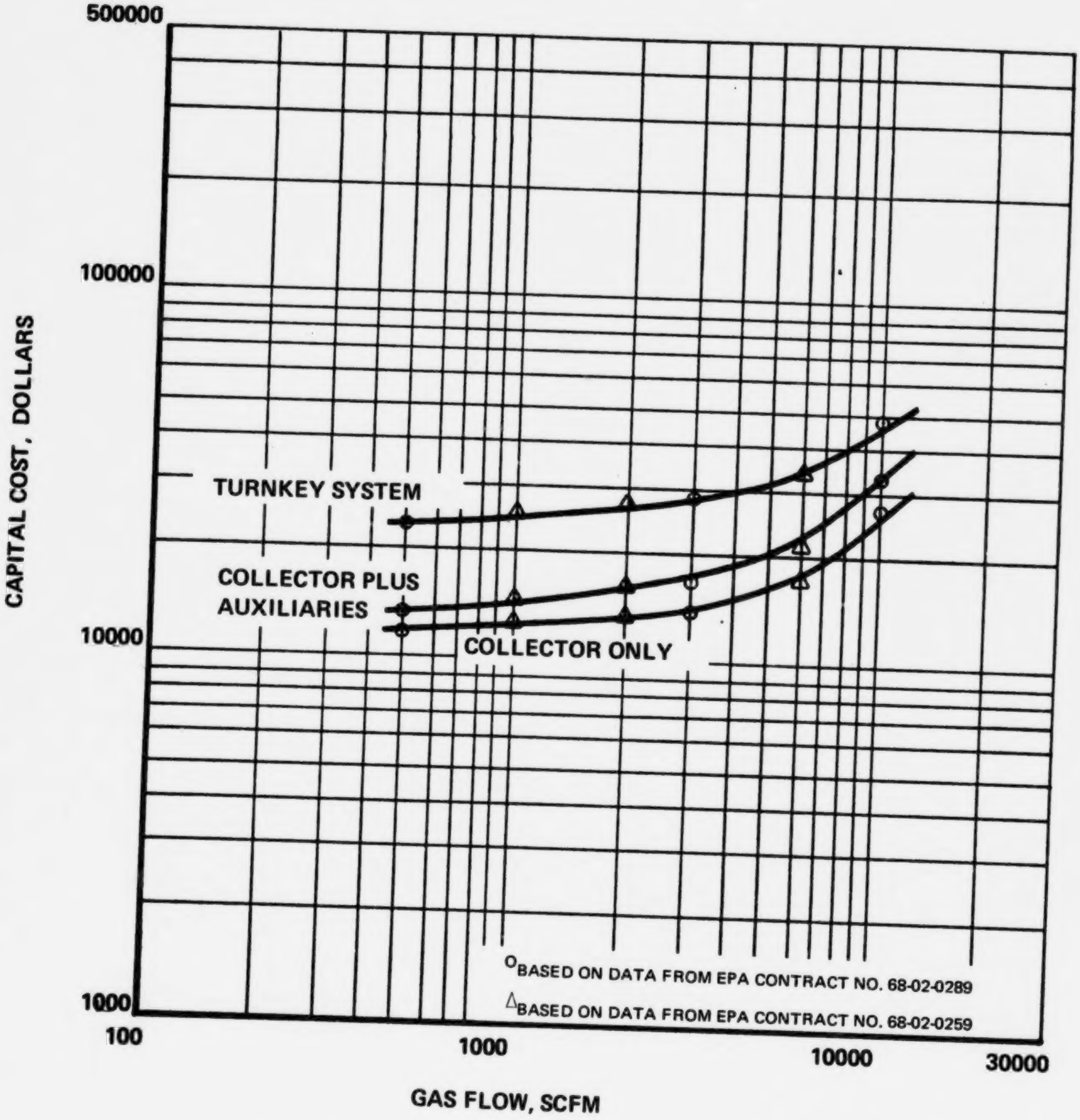
TABLE 108

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR CATALYTIC INCINERATORS FOR OPEN KETTLES (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,760				
Operating Labor (if any)					
Operator	$6/hr			675	675
Supervisor	$8/hr			200	200
Total Operating Labor				875	875
Maintenance					
Labor	$6/hr			510	510
Materials				85	110
Total Maintenance				595	620
Replacement Parts					
Total Replacement Parts				300	1,265
Utilities					
Electric Power	$.011/kw-hr			253	1,118
Fuel	$.80/MM Btu			1,752	10,232
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				2,005	11,350
Total Direct Cost				3,775	14,110
Annualized Capital Charges				2,320	3,875
Total Annual Cost				6,095	17,985

FIGURE 76

CAPITAL COSTS FOR THERMAL INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITHOUT HEAT EXCHANGE)



○ BASED ON DATA FROM EPA CONTRACT NO. 68-02-0289

△ BASED ON DATA FROM EPA CONTRACT NO. 68-02-0259

FIGURE 77

CAPITAL COSTS FOR CATALYTIC INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITHOUT HEAT EXCHANGE)

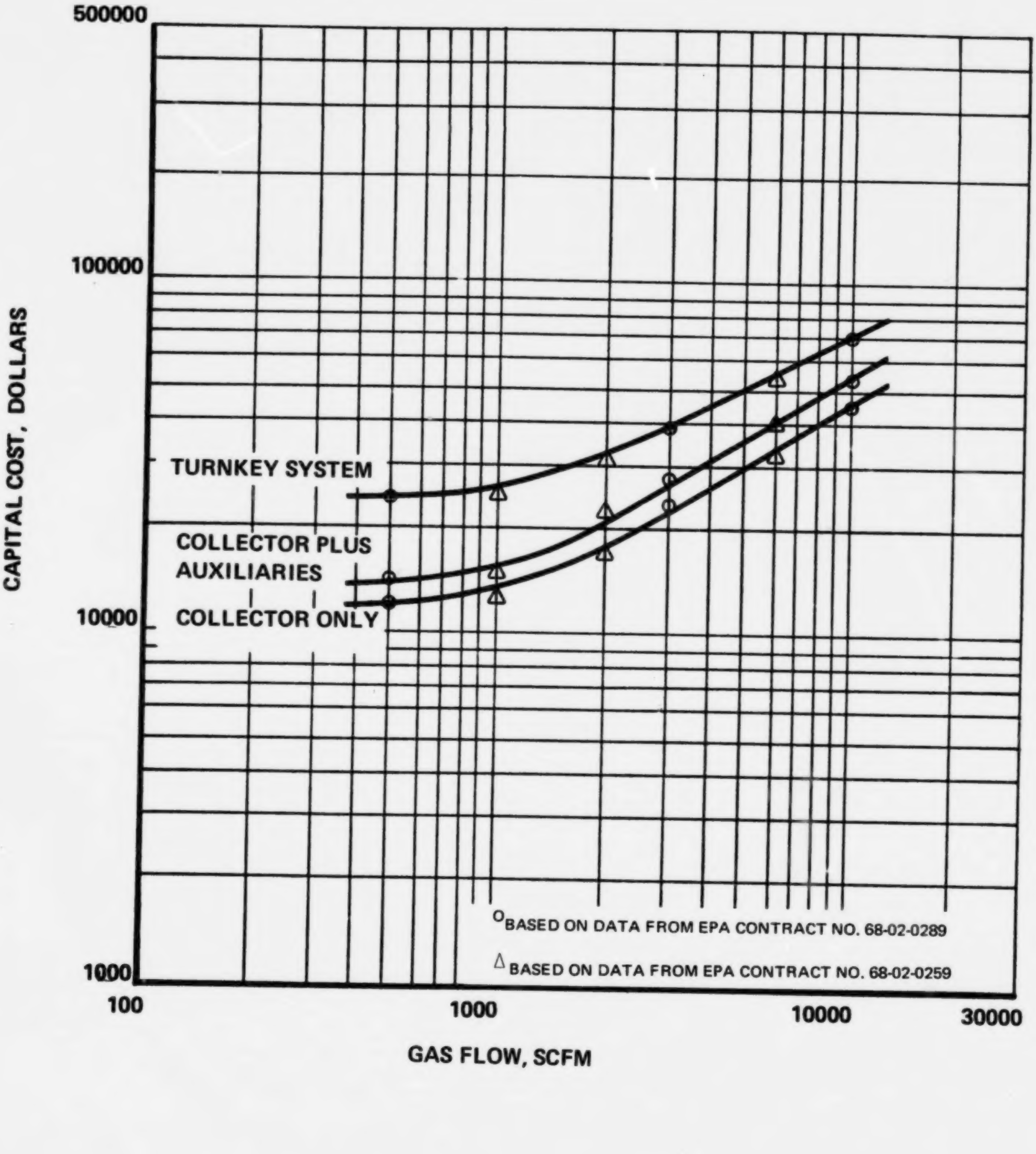


FIGURE 78

DIRECT ANNUAL COSTS FOR THERMAL AND CATALYTIC INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITHOUT HEAT EXCHANGE)

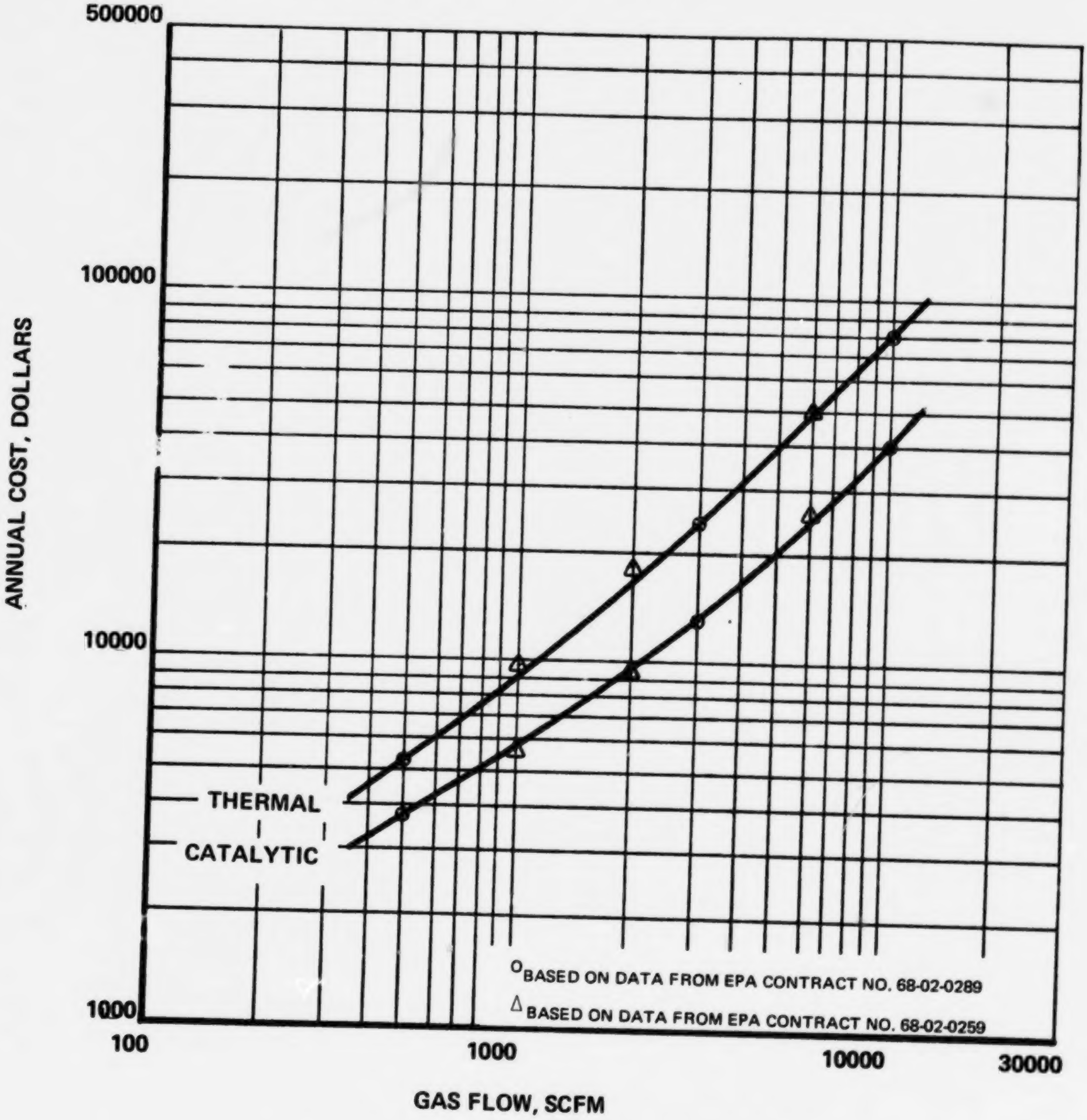


FIGURE 79

TOTAL ANNUAL COSTS FOR THERMAL AND CATALYTIC INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITHOUT HEAT EXCHANGE)

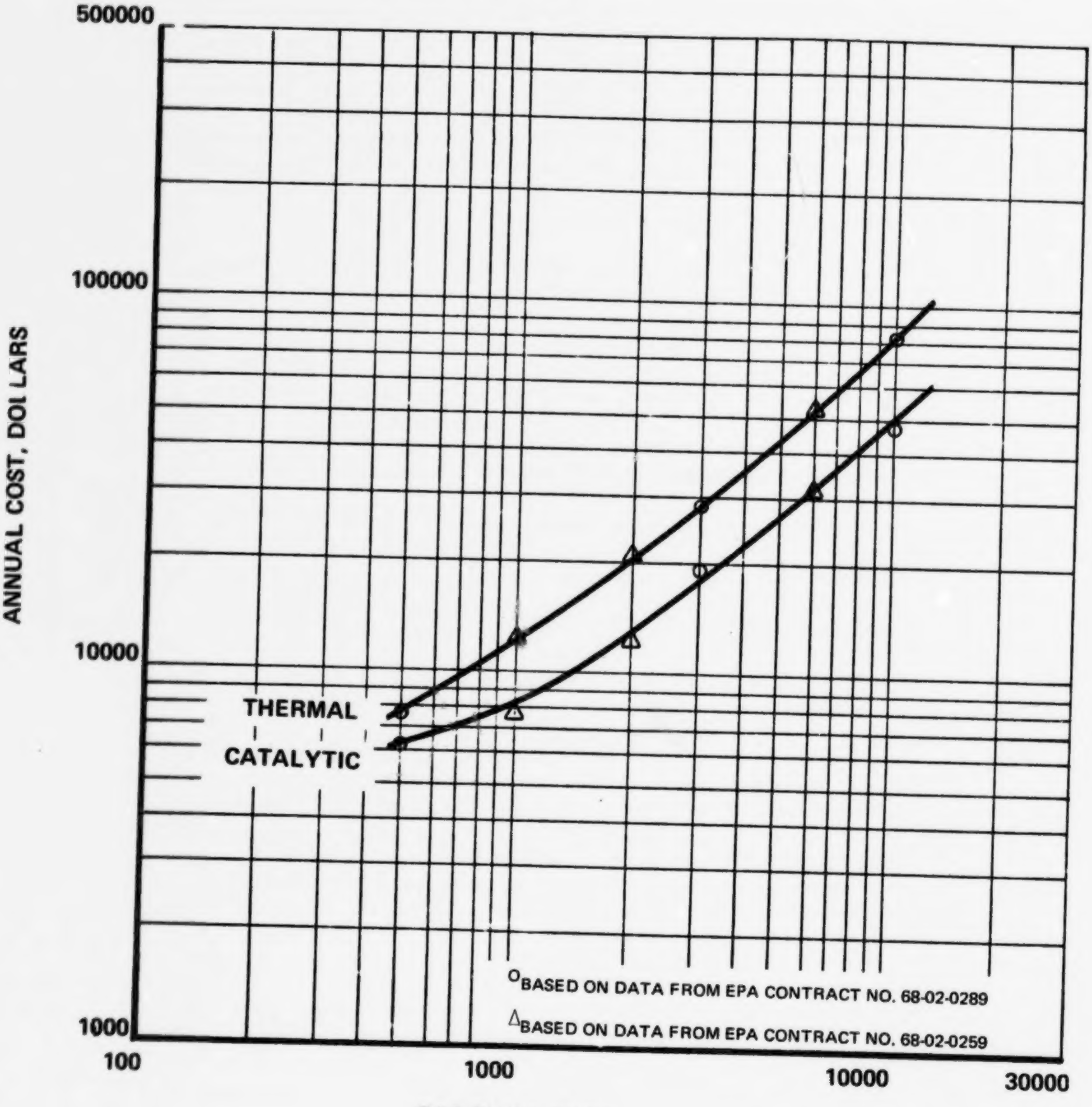


TABLE 109

THERMAL INCINERATOR PROCESS DESCRIPTION FOR RESIN REACTOR SPECIFICATION

This specification describes the requirements for a thermal combustion system for abatement of the hydrocarbon emissions from the resin production facility of a paint and varnish plant. The system will be similar to that shown in Figure 24. All reactor thinning tanks and product rundown tanks will be vented to the collection system. Dilution air will be supplied through a hood over the resin filter press(es). A minimum flow damper is to be supplied in this part of the ventilation system. It is to be sized to allow for a maximum fume concentration of 40% LEL in the fully closed position. A high velocity Venturi section is to be located at the incinerator inlet to assure gas flow is in excess of flame propagation velocity at 1/2 design flow rate.

The afterburner is to be natural gas fired. Sufficient gas, having a specific gravity of 0.60 and an upper heating value of 1040 Btu/SCF, is available at pressure of 1.0 psig.

The exhaust gas contains sufficient oxygen, greater than 16% O_2*, to allow firing of the afterburner with a raw gas or process air burner. A combustion air system is not required. Fume load to the incinerator is composed of 75% kettle emission and 25% from other sources. Average emissions are 60% of peak. The system is to be designed on peak emission but operating costs are to be based on average emissions. Operating conditions listed are based on peak emissions. Fume destruction in burner is to be calculated as follows:*

10% for catalytic units with or without heat exchange

10% for thermal units with heat exchange

20% for thermal units without heat exchange.

Please fill in estimated efficiency of afterburner, burner duty, heat exchanger duty, thermal efficiency, overall heat transfer coefficient (U), and tube surface area.

The afterburner is to be supplied with a suitable control panel and all equipment is to be designed for outdoor operation. Incinerator operating and safety controls are to be designed to meet F.I.A. (Factory Insurance Association) requirements. All dampers are to be pneumatically operated and contain an integral fail-safe air reservoir. The system is to automatically divert in the event of low flow, high afterburner temperature, high reactor pressure, and afterburner preheat burner failure. The exhaust system should also be purged with inert gas on fan failure or loss of flow. Damper operating is to be sequential with the position switches mounted on the damper arm and not the operator. The system fan shall be located after the preheat burner or the incinerator outlet and shall be constructed to withstand 200°F higher than design operating temperature. The fan shall have a V-belt drive and fan and motor shall have the capacity to overcome the pressure drop of the ductwork, afterburner and any heat exchanger that may be used. System ductwork should be sized for a maximum ΔP *of 2 in. w.c. hot. The fan motor may be sized for restricted flow cold start. The ductwork to the incinerator should be heated either by the use of a double manifold or hot gas recycle or a combination of both.*

The heat exchanger is to be the parallel flow shell and tube type and designed to operate at an afterburner outlet temperature of 1500°F. Maximum exchanger pressure drop, both sides, should not exceed a total of 6 in. w.c. hot. Dirty gas is to flow through the tube side.

INSTALLATION

A complete turnkey proposal including ductwork, structural steel, fuel and inert gas piping, etc., is requested. For the purpose of this proposal fan and damper including operators are to be considered as auxiliary. The controls and control cabinet are to be included with the afterburner price. The afterburner will be assumed to be located on a structural steel base on the resin plant roof. No modification to the building structural steel is required. A tie through the roof from the base to the building steel is required. The base and tie-in are part of the installed structural cost. All utilities are available within 30 ft of the control cabinet, motor, and burner. Plant air is available at 100 psig and located within 30 ft. Inert gas is available and will require 30 ft of piping from supply to ductwork. The existing stack can be used for the diversion stack. A 10 ft exhaust stack will be mounted on the incinerator, double manifold or exhaust fan. A total of thirty-five (35) feet of new exhaust ductwork and twenty-five (25) feet of double manifold will be required.

TABLE 110

THERMAL INCINERATOR OPERATING CONDITIONS FOR RESIN REACTOR SPECIFICATION (WITH HEAT EXCHANGE)

	Small	Large
Process Conditions		
Reactors, Number	1	3
Size each, gal	4,000	5,000
Hcbn Emission Max, lb/hr	95	317
Hcbn Emission Ave, lb/hr	57	190
Total Exhaust Rate, SCFM	3,000	10,000
Exhaust Temperature, °F	110	110
Heat of Combustion of Reactor Fume, Btu/lb	17,000	17,000
Hydrocarbon Concentration		
Maximum Btu/SCF	12	12
Average Btu/SCF	7	7
Residence Time, sec @ 1500°F	0.6	0.6
Incinerator with Heat Exchange		
Inlet Tube Side, °F	325	325
Unit Inlet, °F	825	825
Burner ΔT from Fuel Gas, °F	115	115
*Burner ΔT from Flame Combustion, °F	60	60
Burner Outlet Temperature, °F	1,000	1,000
**Unit ΔT from Thermal Combustion, °F	500	500
Unit Outlet Temperature, °F	1,500	1,500
Outlet Shell Side, °F	1,030	1,030
***Burner Duty, MM Btu/hr	1.66	5.4
***H.E. Duty, MM Btu/hr	1.77	5.9
Thermal Efficiency, %	42	42
***Overall Heat Trans Coef, U	4.9	4.9
***Tube Surface Area, ft^2	628	2,095

* Assumes 10% fume combustion in burner flame

** Assumes 95% overall fume combustion

*** Supplied by bidders

Preceding page blank

TABLE 111

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR RESIN REACTORS (WITH HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			3,225	10,750
°F			110	110
SCFM			3,000	10,000
Moisture Content, Vol. %			4.3	4.3
Effluent Contaminant Loading				
lb/hr, Max			95	317
lb/hr, Avg			57	190
Cleaned Gas Flow				
ACFM			8,510	28,400
°F			1,030	1,030
SCFM			3,015	10,050
Moisture Content, Vol. %			7.2	7.2
Cleaned Gas Contaminant Loading				
lb/hr			4.8	16
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			29,674	46,840
(2) Auxiliaries Cost				
(a) Fan(s)			1,863	4,740
(b) Pump(s)				
(c) Damper(s)			1,100	1,405
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			2,050	3,000
(b) Foundations & Support			2,688	4,375
Ductwork			2,600	4,150
Stack			850	1,750
Electrical			600	1,400
Piping			525	720
Insulation				
Painting				
Supervision			900	900
Startup			400	400
Performance Test			1,500	1,500
Other				
(4) Total Cost			44,750	71,180

TABLE 112

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR THERMAL INCINERATORS FOR RESIN REACTORS (WITH HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,760				
Operating Labor (if any)					
Operator	$6/hr			375	375
Supervisor				-	-
Total Operating Labor				375	375
Maintenance					
Labor	$6/hr			390	390
Materials				160	210
Total Maintenance				550	600
Replacement Parts					
Total Replacement Parts				375	475
Utilities					
Electric Power	$.011/kw-hr			2,746	8,528
Fuel	$.80/MM Btu			6,590	21,80C
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				9,336	30,328
Total Direct Cost				10,636	31,778
Annualized Capital Charges				4,475	7,118
Total Annual Cost				15,111	38,896

TABLE 113

CATALYTIC INCINERATOR PROCESS DESCRIPTION FOR RESIN REACTOR SPECIFICATION

This specification describes the requirements for a catalytic combustion system for abatement of the hydrocarbon emissions from the resin production facility of a paint and varnish plant. The system will be similar to that shown in Figure 24. All reactor thinning tanks and product rundown tanks will be vented to the collection system. Dilution air will be supplied through a hood over the resin filter press(es). A minimum flow damper is to be supplied in this part of the ventilation system. It is to be sized to allow for a maximum fume concentration of 40% LEL in the fully closed position. A high velocity Venturi section is to be located at the incinerator inlet to assure gas flow is in excess of flame propagation velocity at 1/2 design flow rate.

The afterburner is to be natural gas fired. Sufficient gas, having a specific gravity of 0.60 and an upper heating value of 1040 Btu/SCF, is available at pressure of 1.0 psig.

The exhaust gas contains sufficient oxygen, greater than 16% O_2, *to allow firing of the afterburner with a raw gas or process air burner. A combustion air system is not required. Fume load to the incinerator is composed of 75% kettle emission and 25% from other sources. Average emissions are 60% of peak. The system is to be designed on peak emission but operating costs are to be based on average emissions. Operating conditions listed are based on peak emissions. Fume destruction in burner is to be calculated as follows:*

10% for catalytic units with or without heat exchange

10% for thermal units with heat exchange

20% for thermal units without heat exchange.

Please fill in estimated efficiency of afterburner, burner duty, heat exchanger duty, overall heat transfer coefficient (U), tube surface area, catalyst face velocity, and catalyst volume.

The afterburner is to be supplied with a suitable control panel and all equipment is to be designed for outdoor operation. Incinerator operating and safety controls are to be designed to meet F.I.A. (Factory Insurance Association) requirements. All dampers are to be pneumatically operated and contain an integral fail-safe air reservoir. The system is to automatically divert in the event of low flow, high afterburner temperature, high reactor pressure, and afterburner preheat burner failure. The exhaust system should also be purged with inert gas on fan failure or loss of flow. Damper operating is to be sequential with the position switches mounted on the damper arm and not the operator. The system fan shall be located after the preheat burner or the incinerator outlet and shall be constructed to withstand 200°F higher than design operating temperature. The fan shall have a V-belt drive and fan and motor shall have the capacity to overcome the pressure drop of the ductwork, afterburner and any heat exchanger that may be used. System ductwork should be sized for a maximum Δ *P of 2 in. w.c. hot. The fan motor may be sized for restricted flow cold start. The ductwork to the incinerator should be heated either by the use of a double manifold or hot gas recycle or a combination of both.*

The heat exchanger is to be the parallel flow shell and tube type and designed to operate at an afterburner outlet temperature of 1500°F. Maximum exchanger pressure drop, both sides, should not exceed a total of 6 in. w.c. hot. Dirty gas is to flow through the tube side.

INSTALLATION

A complete turnkey proposal including ductwork, structural steel, fuel and inert gas piping, etc., is requested. For the purpose of this proposal fan and damper including operators are to be considered as auxiliary. The controls and control cabinet are to be included with the afterburner price. The afterburner will be assumed to be located on a structural steel base on the resin plant roof. No modification to the building structural steel is required. A tie through the roof from the base to the building steel is required. The base and tie-in are part of the installed structural cost. All utilities are available within 30 ft of the control cabinet, motor, and burner. Plant air is available at 100 psig and located within 30 ft. Inert gas is available and will require 30 ft of piping from supply to ductwork. The existing stack can be used for the diversion stack. A 10 ft exhaust stack will be mounted on the incinerator, double manifold or exhaust fan. A total of thirty-five (35) feet of new exhaust ductwork and twenty-five (25) feet of double manifold will be required.

TABLE 114

CATALYTIC INCINERATOR OPERATING CONDITIONS FOR RESIN REACTOR SPECIFICATION (WITH HEAT EXCHANGE)

	Small	Large
Process Conditions		
Reactor Number	1	3
Size each, gal	4,000	5,000
Hcbn Emission Max, lb/hr	95	317
Hcbn Emission Ave, lb/hr	57	190
Total Exhaust Rate, SCFM	3,000	10,000
Exhaust Temperature, °F	110	110
Heat of Combustion of Resin Fume, Btu/lb	17,000	17,000
Hydrocarbon Concentration		
Maximum Btu/SCF	12	12
Average Btu/SCF	7	7
Incinerator with Heat Exchange		
Inlet Tube Side, °F	300	300
Unit Inlet, °F	500	500
Burner ΔT from Fuel Gas, °F	40	40
*Burner ΔT from Flame Combustion, °F	60	60
Burner Outlet Temperature, °F	600	600
**Unit ΔT from Thermal Combustion, °F	500	500
Unit Outlet Temperature, °F	1,100	1,100
Outlet Shell Side, °F	915	915
Burner Duty, MM Btu/hr	– – –	– – –
HE Duty, MM Btu/hr	– – –	– – –
Thermal Efficiency	~ 23%	~ 23%
***Overall Heat Trans Coef, U	4.2	4.2
***Tube Surface Area, ft^2	405	1,350

* Assumes 10% fume combustion in burner flame
** Assumes 95% overall fume combustion
*** Supplied by bidders

Preceding page blank

TABLE 115

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CATALYTIC INCINERATORS FOR RESIN REACTORS (WITH HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			3,225	10,750
°F			110	110
SCFM			3,000	10,000
Moisture Content, Vol. %			4.3	4.3
Effluent Contaminant Loading				
lb/hr, Max			95	317
lb/hr, Avg			57	190
Cleaned Gas Flow				
ACFM			7,775	25,900
°F			915	915
SCFM			3,002	10,007
Moisture Content, Vol. %			7.0	7.0
Cleaned Gas Contaminant Loading				
lb/hr			4.8	15.9
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			31,698	53,890
(2) Auxiliaries Cost (a) Fan(s) (b) Pump(s) (c) Damper(s) (d) Conditioning, Equipment (e) Dust Disposal Equipment			4,608	10,230
(3) Installation Cost (a) Engineering (b) Foundations & Support Ductwork Stack Electrical Piping Insulation Painting Supervision Startup Performance Test Other			11,561	17,173
(4) Total Cost			47,867	81,293

TABLE 116

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR CATALYTIC INCINERATORS FOR RESIN REACTORS (WITH HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year	8,760				
Operating Labor (if any)					
Operator	$6/hr			675	675
Supervisor	$8/hr			200	200
Total Operating Labor				875	875
Maintenance					
Labor	$6/hr			510	510
Materials				115	145
Total Maintenance				625	655
Replacement Parts					
Total Replacement Parts				1,265	4,045
Utilities					
Electric Power	$.011/kw-hr			2,004	5,436
Fuel	$.80/MM Btu			841	3,083
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				2,845	8,519
Total Direct Cost				5,610	14,094
Annualized Capital Charges				4,787	8,129
Total Annual Cost				10,397	22,223

FIGURE 80

CAPITAL COSTS FOR THERMAL INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITH 42% EFFICIENT HEAT EXCHANGE)

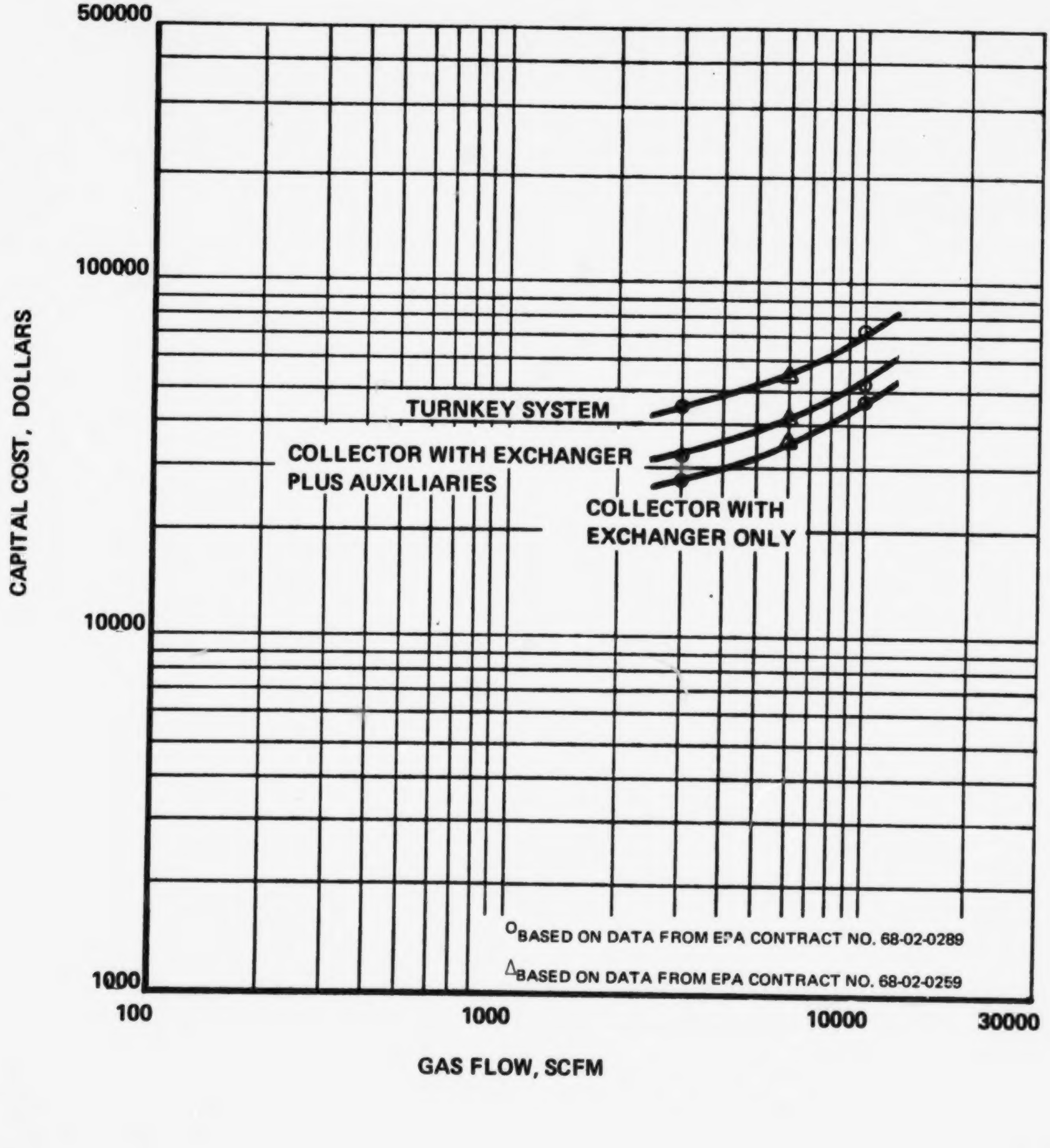


FIGURE 81

CAPITAL COSTS FOR CATALYTIC INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITH 23% EFFICIENT HEAT EXCHANGE)

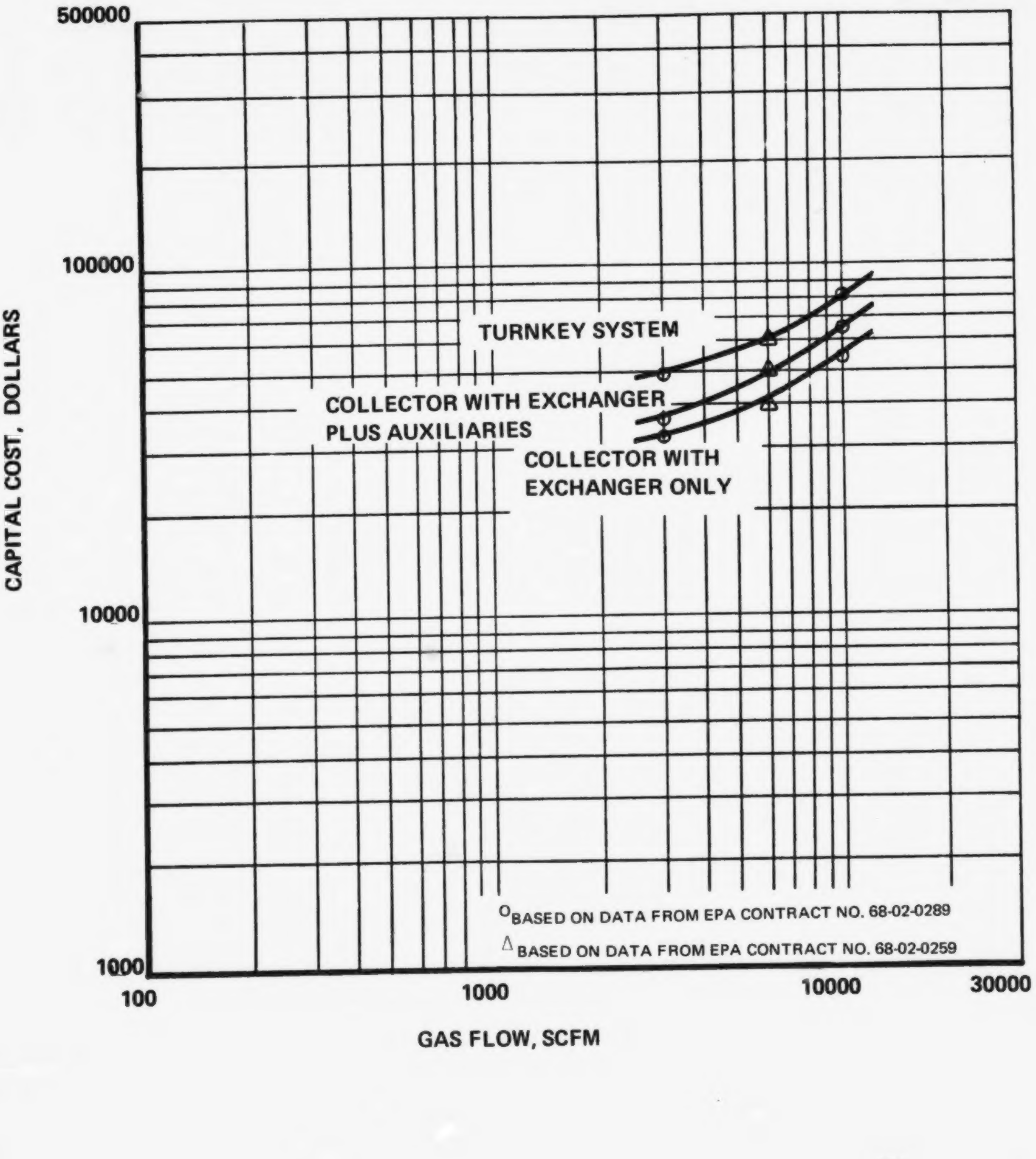


FIGURE 82

DIRECT ANNUAL COSTS FOR THERMAL AND CATALYTIC INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITH HEAT EXCHANGE)

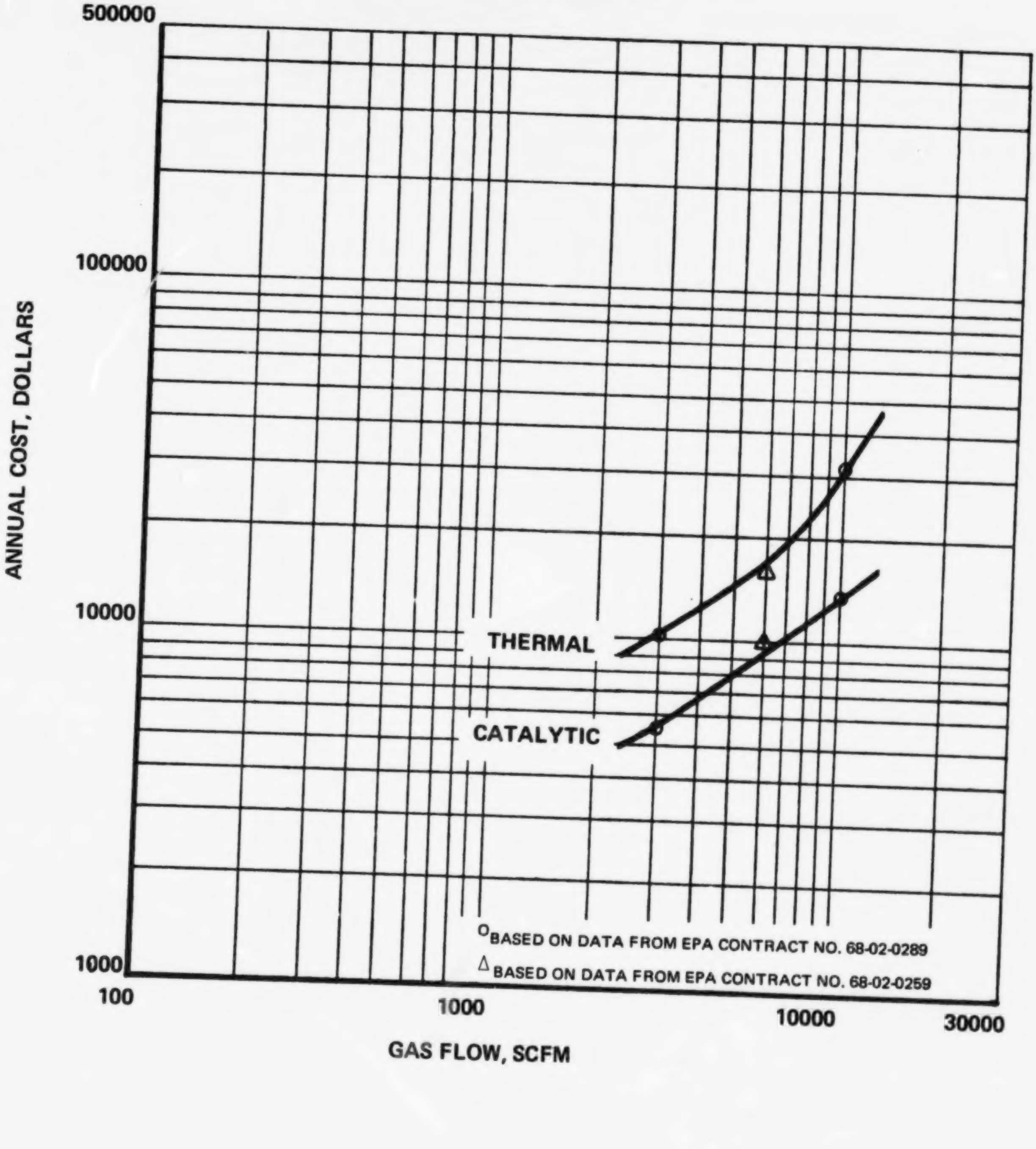


FIGURE 83

TOTAL ANNUAL COSTS FOR THERMAL AND CATALYTIC INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY (WITH HEAT EXCHANGE)

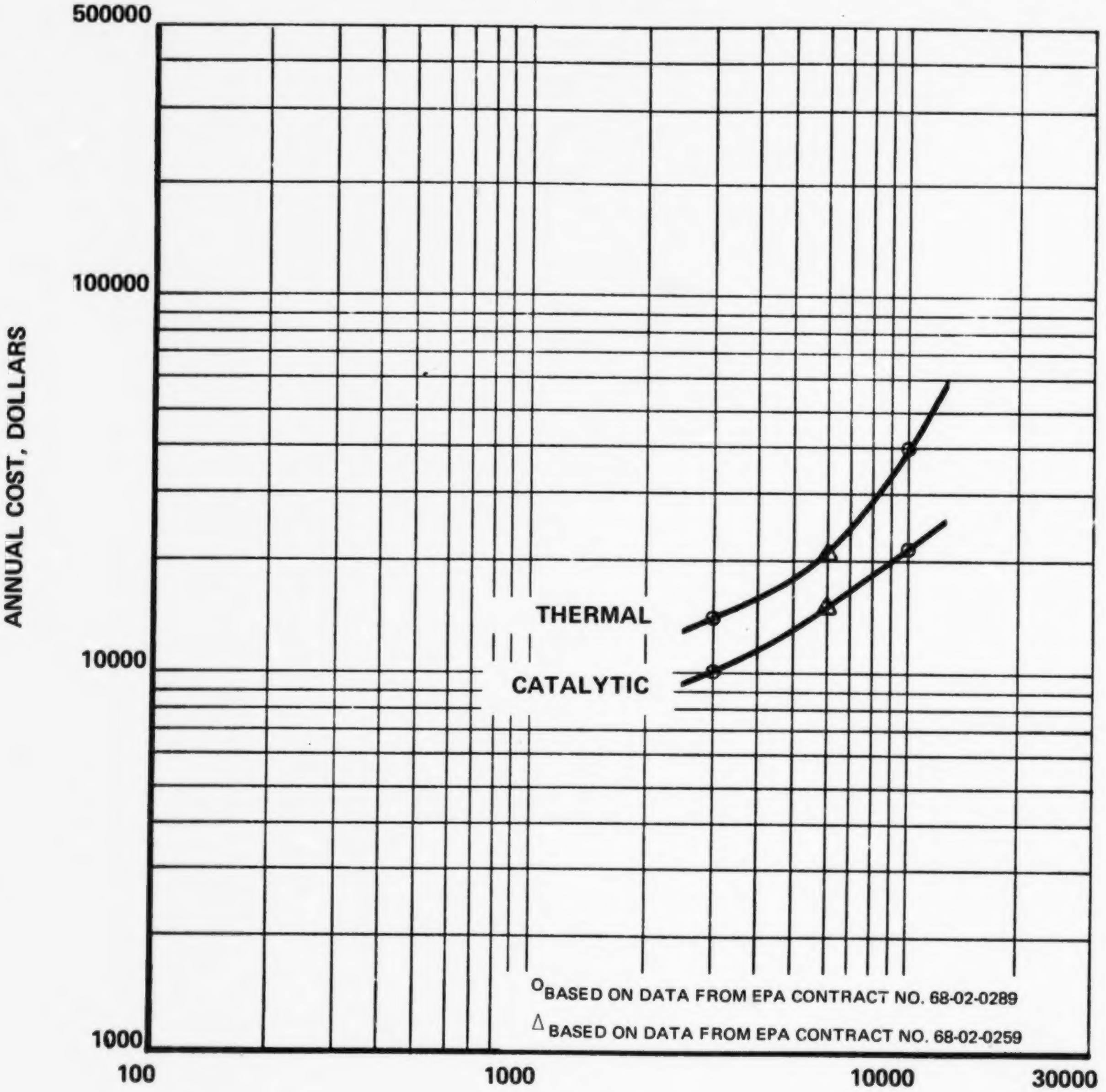
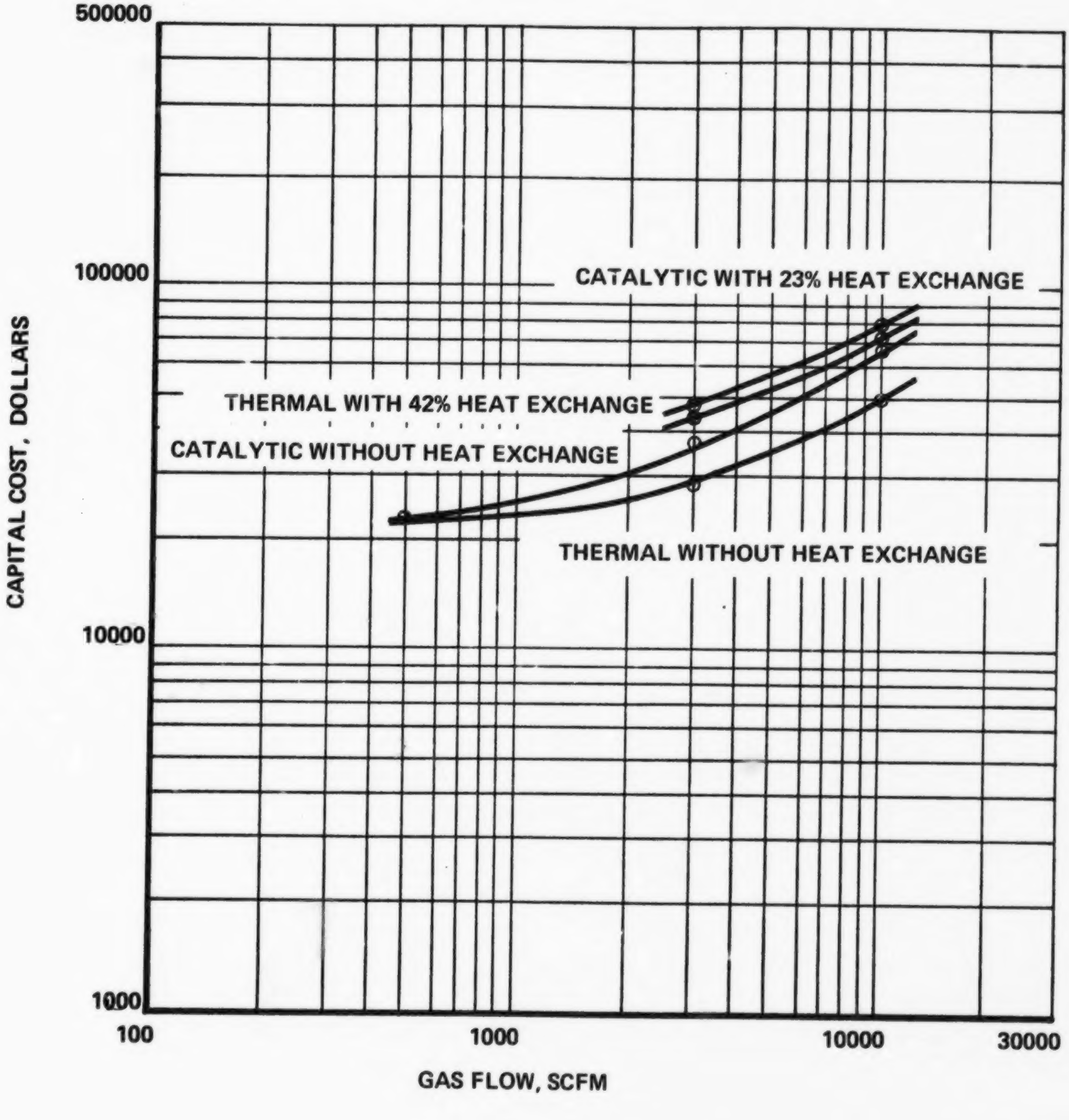


FIGURE 84

TOTAL INSTALLED COSTS FOR THERMAL AND CATALYTIC INCINERATORS FOR THE PAINT AND VARNISH INDUSTRY



Preceding page blank

REFERENCES

1. Marketing Guide to the Paint Industry, Patricia Noble, Ed., Charles H. Kline & Co., Inc., Fairfield, New Jersey, 1969.

2. Federation Series on Coating Technology, Units
   1. Introduction to Coating Technology, Oct., 1964
   3. Oils for Organic Coatings, Sept., 1965
   4. Modern Varnish Technology, May, 1966
   5. Alkyd Resins, March, 1967
   12. Principles of Formulation and Paint Calculation, June, 1969
   17. Acrylic Resins, March, 1971
   19. Vinyl Resins, April, 1972, Federation of Societies for Paint Technology, Philadelphia, Pennsylvania

3. Stenburg, R. L., Control of Atmospheric Emissions from Paint and Varnish Manufacturing Operations, U.S. Department of Health, Education and Welfare, Robert A. Taft Sanitary Engineering Center, Technical Report A58-4, Cincinnati, Ohio (June, 1958).

4. Parker, Dean H., Principles of Surface Coating Technology, John Wiley & Sons, Interscience Publishers Division, New York, N.Y., 1965.

5. The Technology of Paints, Varnishes and Lacquers, Charles R. Martens, Ed., Reinhold Book Corp., New York, N.Y., 1968.

6. Spence, J. W. and Haynie, F. H., Paint Technology and Air Pollution: A Survey and Economic Assessment, Environmental Protection Agency, National Environmental Research Center, Office of Air Programs Publication No. AP-103, Research Triangle Park, North Carolina, February, 1972.

7. Hardison, L. C., "A Summary of the Use of Catalyst for Stationary Emission Source Control", presented at Franklin Institute, Philadelphia, Pennsylvania (Nov., 1968).

8. Hardison, L. C., "Disposal of Gaseous Wastes", presented at East Ohio Gas Company Seminar on Waste Disposal, Cleveland, Ohio (May, 1967).

9. Hardison, L. C., "Controlling Combustible Emissions", Paint and Varnish Production, July, 1967.

10. Unpublished source tests of 11 different kettles, Hirt Combustion Engineers, Montebello, California, August, 1971.

11. Unpublished source tests of 10 different kettles, PPG Industries.

12. Rolke, R. W., et al, Afterburner Systems Study, Shell Development Co., Emeryville, California.

4. THE GRAPHIC ARTS INDUSTRY

The graphic arts industry encompasses all operations involved in the printing of an image onto a surface. The images may consist of alphabetical and numerical figures, illustrations, and photographs, while the surfaces may consist of various grades of paper, metal, plastic, fabric, and glass. The industry is composed of over 40,000 commercial establishments(1) scattered throughout the country and ranks seventh in value-added to the Gross National Product and eighth in employment.(2) Although the industry is large on a national scale, it is one of the few remaining industries composed mainly of small businesses. Approximately 80%(3) of the commercial printing establishments employ less than twenty people.

The industry can be classified into three general categories:

1. Publishing – newspapers, books, magazines, etc.

2. Package Printing – paper containers, labels, etc.

3. Metal Decorating – metal containers, bottle caps, metal signs, etc.

Generally, these categories differ in the type of surfaces (substrates) to which the printing images are applied. The publishing and package printing industries generally employ various grades of paper as substrates, while the metal decorating industry employs various types of sheet metal.

The majority of commercial printing operations employ one of the following printing processes:

Letterpress
Lithography
Gravure
Flexography
Screen Printing

Each process has a unique method of applying the printing image onto the substrate and a unique ink composition.

In letterpress printing, the printing image* is relief outward from the non-image area. Originally, the printing surface was a flat plate; however, in modern letterpress printing, the printing image is transferred to a flexible mat from which a cylindrical plate is made. The curved plates are then attached to

---

*That portion of the printing plate or cylinder which transfers the ink to the substrate.

Preceding page blank

rollers in the press. The process is capable of both sheet-fed and web-fed (roll) printing with press speeds up to 1500 feet per minute (web).

Letterpress is the oldest and most basic form of printing and still predominates in periodical and newspaper publishing. Approximately 93% of the nation's newspapers are printed by this process.(4) Letterpress accounts for 35.6% of the eight billion dollar commercial printing market (Table 117)(5) and 57.3% of the industry's ink consumption (Table 118).(6)

Lithography printing is characterized by having the image area on the same plane as the non-image area. The image area chemically attracts ink (ink receptive area) while the non-image area chemically repels ink (water receptive area). The printing image is applied to a cylinder which transfers the inked image directly to the substrate (direct lithography) or to a rubber blanket cylinder which in the same revolution prints the wet inked image onto the substrate (offset lithography). When a web or continuous roll of paper is employed with the offset process, it is called web-offset printing.

Lithography currently accounts for 50.9% of the commercial printing market and 10.9% of the industry's ink consumption. It is considered the fastest growing printing process and finds its greatest use in books, pamphlets, metal decorating, and newspapers of over 100,000 circulation.

In gravure or intaglio printing, the image area is relief inward from the non-image area. During the inking process, both the image and non-image surfaces are submerged in an ink bath, and the excess ink is scraped from the non-image area with a blade (Doctor's blade). After removal of the excess ink, the inked image area is transferred directly to the substrate. In sheet-fed gravure, the image carrier may be either a flat plate or a curved plate attached to a cylinder. In roll-fed gravure or rotogravure, the image is engraved into a metal cylinder. Rotogravure presses are capable of press speeds up to 2,000 feet per minute.(2)

Gravure printing accounts for 6.6% of the commercial printing market and 19.1% of the industry's ink consumption. The process is widely used in package printing (cellophane, metallic foil, gift wrap, and labels), magazines, catalogues, and Sunday newspaper supplements.

Flexography is very similar to letterpress in that the printing image is relief outward from the non-image area; however, in flexography the printing plate is made of rubber. The process is always web-fed and is used primarily in package printing (milk cartons).

TABLE 117

COMMERCIAL PRINTING MARKET(5)

Printing Process	Value of Receipts ($ Million) 1963	1967	1970	Percentage of Market 1963	1967	1970
Letterpress	2,074	2,236	2,814	45.3	38.2	35.6
Lithography	1,896	2,787	4,032	41.3	47.5	50.9
Gravure	287	207	525	6.2	6.9	6.6
Flexography	263	304	337	5.7	5.2	4.3
Screen Printing	69	130	210	1.5	2.2	2.6
Total	4,589	5,664	7,918	100.0	100.0	100.0

TABLE 118

INK* CONSUMPTION IN 1968(6)

Printing Process	Ink* Consumed Pounds x 10^{-6}	Ink* Consumed % of Total
Lithography	81.5	11.0
Letterpress	459.0	57.0
(Publication)	(143.0)	(18.0)
(Newspaper)	(316.0)	(39.0)
Gravure	153.0	18.0
Flexography	66.5	9.0
Screen Printing	41.0	5.0
Total	801.0	100.0

*Undiluted Ink

In screen printing, the printing image is a fine screen through which the ink or paint flows. Non-image areas are produced by impregnating the screen with a waxy material which prevents flow of the ink through the screen. This process is entirely sheet-fed and is employed mainly in the printing of greeting cards, signs, and wallpaper.

INKS AND SOLVENTS

In the printing processes, the main sources of air pollution result from the printing inks which usually consist of three components.

1. Pigments, which produce the desired colors, are composed of finely divided organic and inorganic materials.
2. Resins, which bind the pigments to the substrate, are composed of organic resins and polymers.
3. Solvents, which dissolve or disperse the resins and pigments, are usually composed of organic compounds. The solvent is removed from the ink and emitted to the atmosphere during the drying process.

Solvents can be classified into five general categories according to their chemical composition.

1. Aromatics: benzene, toluene, xylene, ethylbenzene, unsaturates, and mixtures with aromatic contents greater than 25%.
2. Aliphatics and Intermediates: normal and isoparaffins, cycloparaffins, and mineral spirits containing less than 15% aromatics.
3. Oxygen-containing Compounds: methanol, propanol, isopropanol, butanol, isobutanol, glycols, esters, and ketones.
4. Chlorinated Compounds: trichloroethylene, trichloroethane, methylene chloride.
5. Nitrogen-containing Compounds: nitroparaffins and dimethyl formamide.

The distribution of solvent usage among the various printing processes is tabulated in Tables 120[2] and 121[2] and chemical properties of representative solvents for each group are contained in Table 119.[1]

The viscosity of printing inks is controlled by the solvent content and may vary from a paste-like substance in letterpress and lithography to a very fluid substance in gravure. Gravure inks must be fluid enough to flow into and out of the engraved cells of the printing cylinder, while letterpress and lithography inks must be viscous enough to adhere to the image areas during the printing process.

Gravure inks usually have a solvent concentration of approximately 65% and are composed of highly volatile aromatics (toluene and xylene) which evaporate very rapidly at room temperature. Letterpress and lithography inks can be divided into two categories: oxidative drying inks and heat set inks. The oxidative drying inks (newspaper printing) contain very little solvent and dry by oxidative polymerization without the addition of heat. Heatset inks contain approximately 40% aliphatic solvent, and during the high temperature (≃ 400°F) drying operation, approximately 50% of the original solvent is removed and emitted to the atmosphere.

Small quantities of solvents are also consumed in the printing industry as washing solutions and varnish thinners. The relative quantities consumed for these purposes are tabulated in Tables 122[2] and 123.[2]

SOLVENT FLOW DIAGRAMS AND EMISSION RATES

Web-offset – Heatset Ink

A solvent flow diagram of a web-offset operation employing a heatset ink is illustrated in Figure 85. The web travels through the presses (one for each color), where it is printed on both sides simultaneously. The wet web is then passed through a dryer (≃ 400°F) where approximately 60% of the initial solvent is removed, and then through chill rolls where it is cooled prior to folding and cutting.

The dryer may be either a hot air dryer (as shown in the illustration) where a minimum of flame impingement occurs or an all flame dryer where direct impingement of the flame on the web occurs. The composition of the dryer effluent gas depends on the type of dryer being employed.

In the hot air dryer, very little if any solvent decomposition occurs. As the

TABLE 119

PHYSICAL PROPERTIES OF COMMONLY USED SOLVENTS[1]

SOLVENT	BOILING RANGE, °F	FLASH POINT, °F	lb/gal @ 68°F
ESTERS:			
Methyl Acetate	125 to 136	less than 30	7.55
Ethyl Acetate	161 to 176	42	7.40
Isopropyl Acetate	183 to 194	60	7.25
Secondary Butyl Acetate	219 to 242	90	7.18
Normal Butyl Acetate	239 to 266	102	7.30
Amyl Acetate	248 to 302	116	7.20
ALCOHOLS:			
Methyl Alcohol	147 to 149	61	6.60
Isopropyl Alcohol	177 to 181	70	6.55
Normal Butyl Alcohol	240 to 246	115	6.75
GLYCOL - ETHER:			
Methyl "Cellosolve"	253 to 258	115	8.03
"Cellosolve"	269 to 278	130	7.74
KETONES:			
Acetone	132 to 134	less than 15	6.60
Methyl Ethyl Ketone	172 to 177	less than 30	6.71
Methyl Isobutyl Ketone	237 to 242	80	6.68
Cyclohexanone	266 to 343	130	7.88
AROMATICS:			
Toluene	229 to 233	45	7.24
Xylene	275 to 290	82	7.26
ALIPHATIC NAPHTHAS:			
Hexane	150 to 158	less than 0	5.70
Fast Diluent Naphtha	140 to 180	less than 0	5.75
Heptane	200 to 210	less than 0	6.05
Lacquer Diluent Naphtha	200 to 240	less than 30	6.20
Octane	215 to 230	30	6.20
Mineral Spirits	310 to 400	100	6.50

TABLE 120

PERCENTAGE BREAKDOWN OF SOLVENT CONSUMED FOR INK DILUTION BY PRINTING PROCESS AND SOLVENT TYPE (1968)[2]

PRINTING PROCESS	SOLVENT TYPE (VOLUME %) A	B	C	D	E	F	TOTAL
Lithography	12.8	20.4	14.4	0.04	0.58	0.30	48.52%
Letterpress	0.1	0.4	0.3	0.06	–	–	0.86%
Flexography	0.05	0.5	8.7	–	–	0.15	9.40%
Gravure	8.6	21.1	11.1	–	0.02	–	40.82%
Screen Printing	0.05	0.2	–	–	–	0.15	0.40%
Total	21.60%	42.60%	34.50%	0.10%	0.60%	0.60%	100.0%*

*Total solvent consumed as ink diluent was 11,682,500 gal in 1968.

Solvent Types:

A – Benzene, toluene, xylene, ethylbenzene, unsaturates and mixtures with aromatic content greater than 25% by volume

B – Normal and isoparaffins, cycloparaffins, mineral spirits containing less than 15% aromatics, heavy naphthas with aromatic contents greater than 25%, turpentine

C – Methanol, ethanol, propanol, isopropanol, butanol, isobutanol, glycols, esters, ketones

D – Trichloroethylene, trichloroethane, methylene chloride

E – Nitroparaffins and dimethyl formamide

F – Miscellaneous

TABLE 121

VOLUME BREAKDOWN OF SOLVENT CONSUMED FOR INK DILUTION BY PRINTING PROCESS AND SOLVENT TYPE (1968)(2)

PRINTING PROCESS	SOLVENT TYPE (HUNDRED GALLONS) A	B	C	D	E	F	TOTAL
Lithography	14,972	23,941	16,691	38	723	408	56,773
Letterpress	98	444	399	52	–	1	994
Flexography	58	606	10,180	–	1	170	11,015
Gravure	10,089	24,637	12,868	–	12	–	47,606
Screen Printing	34	173	85	–	–	145	437
Total	25,251	49,801	40,223	90	736	724	116,825

Solvent Type – See Table 60

TABLE 122

VOLUME BREAKDOWN OF TOTAL SOLVENT CONSUMPTION BY USE AND SOLVENT TYPE (1968)[2]

USE	SOLVENT TYPE (THOUSAND GALLONS)				
	A	B	C	D, E, & F	TOTAL
Ink Dilution	2,525	4,980	4,022	155	11,682
Washes*	579	1,774	519	–	2,872
Coating & Varnish Thinner	614	1,156	984	–	2,754
Total	3,718	7,910	5,525	155	17,308

*Washing presses, rollers, blankets, etc.

Solvent Type – See Table 60

TABLE 123

PERCENTAGE BREAKDOWN OF TOTAL SOLVENT CONSUMPTION BY USE AND SOLVENT TYPE (1968)(2)

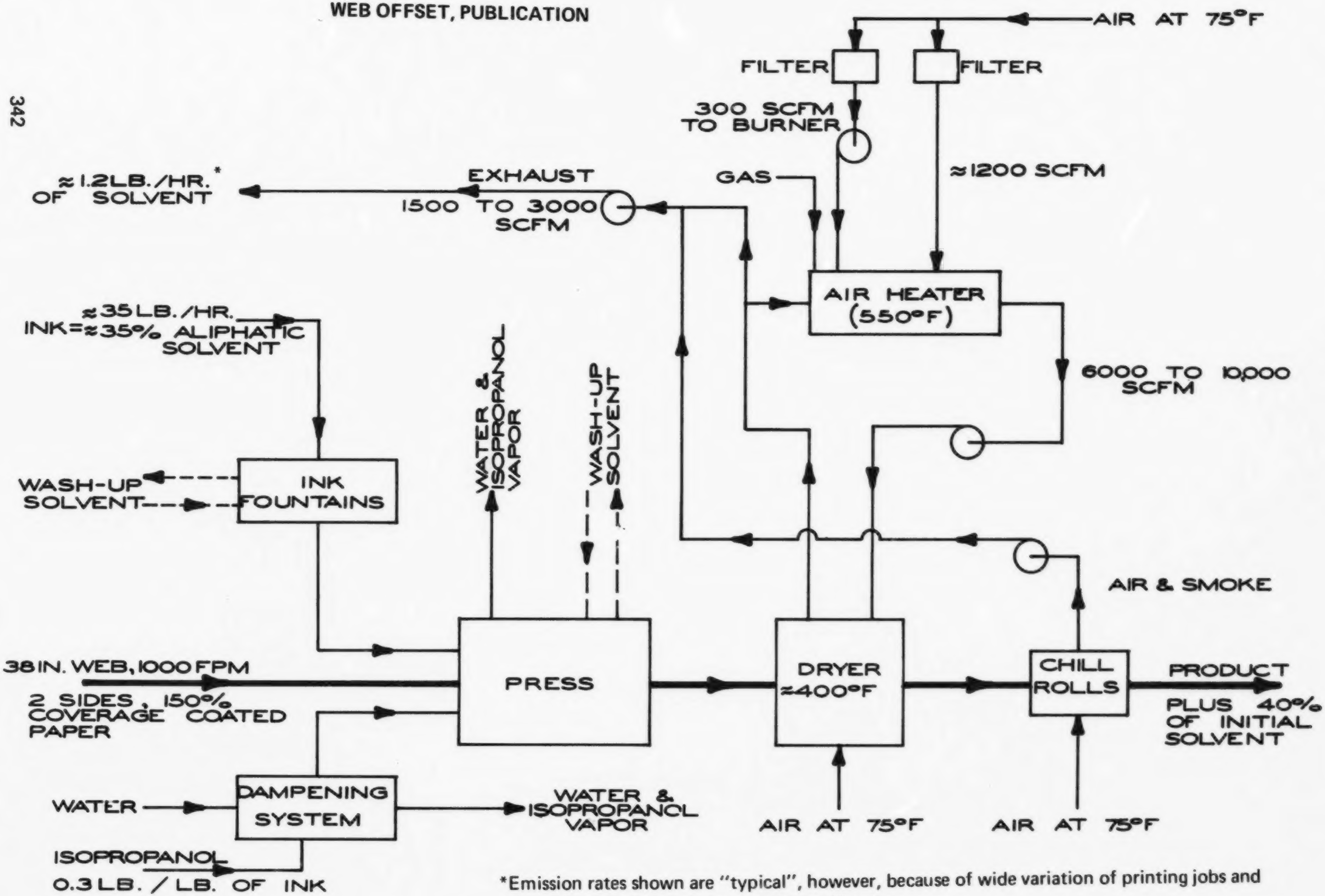
USE	A	B	C	D. E, & F	TOTAL
Ink Dilution	14.5	28.8	23.2	1.0	67.5
Washes*	3.4	10.2	3.0	–	16.6
Coating & Varnish Thinner	3.5	6.7	5.7	–	15.9
Total	21.4	45.7	31.9	1.0	100.0%

*Washing presses, rollers, blankets, etc.

Solvent Type – See Table 60

FIGURE 85

WEB OFFSET, PUBLICATION


AIR AT 75°F
FILTER
FILTER
300 SCFM TO BURNER
GAS
≈1200 SCFM
AIR HEATER (550°F)
6000 TO 10,000 SCFM
EXHAUST
1500 TO 3000 SCFM
≈1.2LB./HR.* OF SOLVENT
≈35LB./HR. INK=≈35% ALIPHATIC SOLVENT
WASH-UP SOLVENT
INK FOUNTAINS
WATER & ISOPROPANOL VAPOR
WASH-UP SOLVENT
38IN. WEB, 1000FPM
2 SIDES, 150% COVERAGE COATED PAPER
PRESS
DRYER ≈400°F
CHILL ROLLS
AIR & SMOKE
PRODUCT
PLUS 40% OF INITIAL SOLVENT
WATER
DAMPENING SYSTEM
WATER & ISOPROPANOL VAPOR
ISOPROPANOL
0.3LB./LB. OF INK
AIR AT 75°F
AIR AT 75°F

*Emission rates shown are "typical", however, because of wide variation of printing jobs and operating conditions, large variations will exist.

amount of flame impingement increases, the quantity of solvent decomposition also increases.

Exhaust gas analyses from a web-offset publication operation* with and without ink and for coated and non-coated paper are included in Table 124.(2) These analyses indicate that the ink is not the only source of emissions from the printing operation. The paper coating also emits substantial quantities of hydrocarbons.

The operation illustrated in Figure 85 represents a "typical" web-offset, heatset ink, printing process operating at the following conditions:

Web Width	38 in.
Press Speed	1,000 ft/min
Number of Sides Printed	2
Ink Coverage	150%
Ink Mileage	0.41 lb ink/10^6 in.^{2}
Solvent Content of Ink	35% (Aliphatic)
% of Initial Solvent Removed in Dryer	60%

The emission rate of solvent from the operation as a function of press speed is shown in Figure 86.

Letterpress – Publication

Letterpress differs from lithography (web-offset) in that the web is printed on one side at a time, and the web is dried after each color is printed. When four colors are printed, a procedure called double ending may be employed: the web passes through one press and one dryer, is then turned over, and returned to the same press where it runs adjacent to the first pass on the same cylinder. In this manner, only four presses and four dryers are required for four-color, two-sided printing. The exhaust and solvent emission rates shown in the solvent flow diagram (Figure 87) represent one color, two-sided printing. In an actual four-color operation, the four dryers would be manifolded together to a common exhaust stack and the emission rates would be four times as great as those shown in the illustration.

Letterpress publication ink is very similar in composition to the web-offset ink (heatset – 35% aliphatic solvent), and consequently, the

---

*Dryer type not included.

TABLE 124

ANALYSIS OF DRYER EXHAUST GAS FROM WEB-OFFSET PUBLICATION(2)

Component	Web-Offset – Publication			
	A	B	C	D
CO, ppm	55	36.5	7.3	55
CH_4, ppm	18	9	9.1	18
CO_2, ppm	3,010	2,970	3,180	2,850
HCBN*	1.03	0.703	0.297	0.985
Carbon, lb/hr	0.279	0.190	0.077	0.266

*CH_4 free hydrocarbons, (lb C/ft^3) x 10^5

A – Coated paper, 30 in. web, 60% coverage, 4 color, one side 1,500 impressions/min, 300° F oven temperature, exhaust = 4,500 SCFM

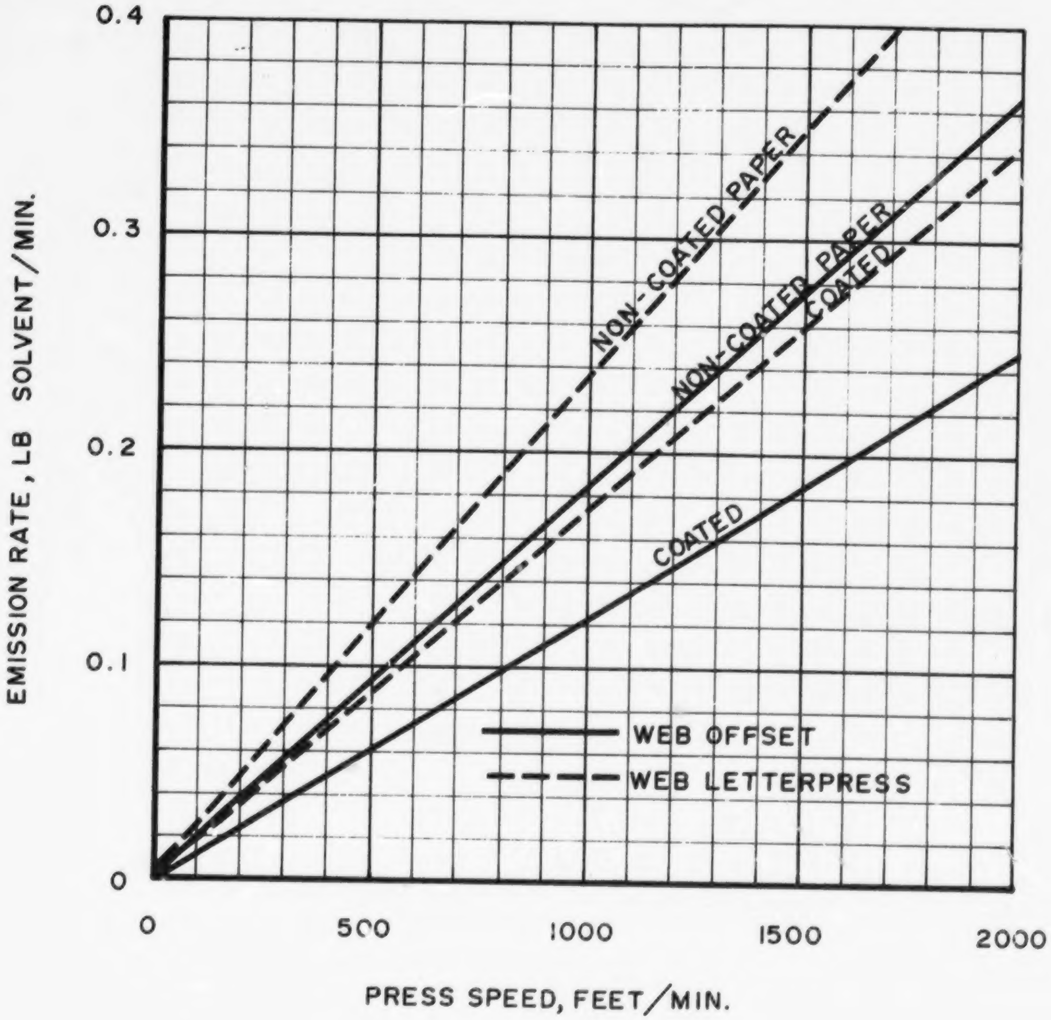
B – Same as A except no ink was used.

C – News stock (uncoated) paper, same conditions as B. No ink.

D – News stock, same conditions as A.

FIGURE 86

EMISSION RATES* FROM WEB OFFSET AND WEB LETTERPRESS EMPLOYING HEATSET INKS

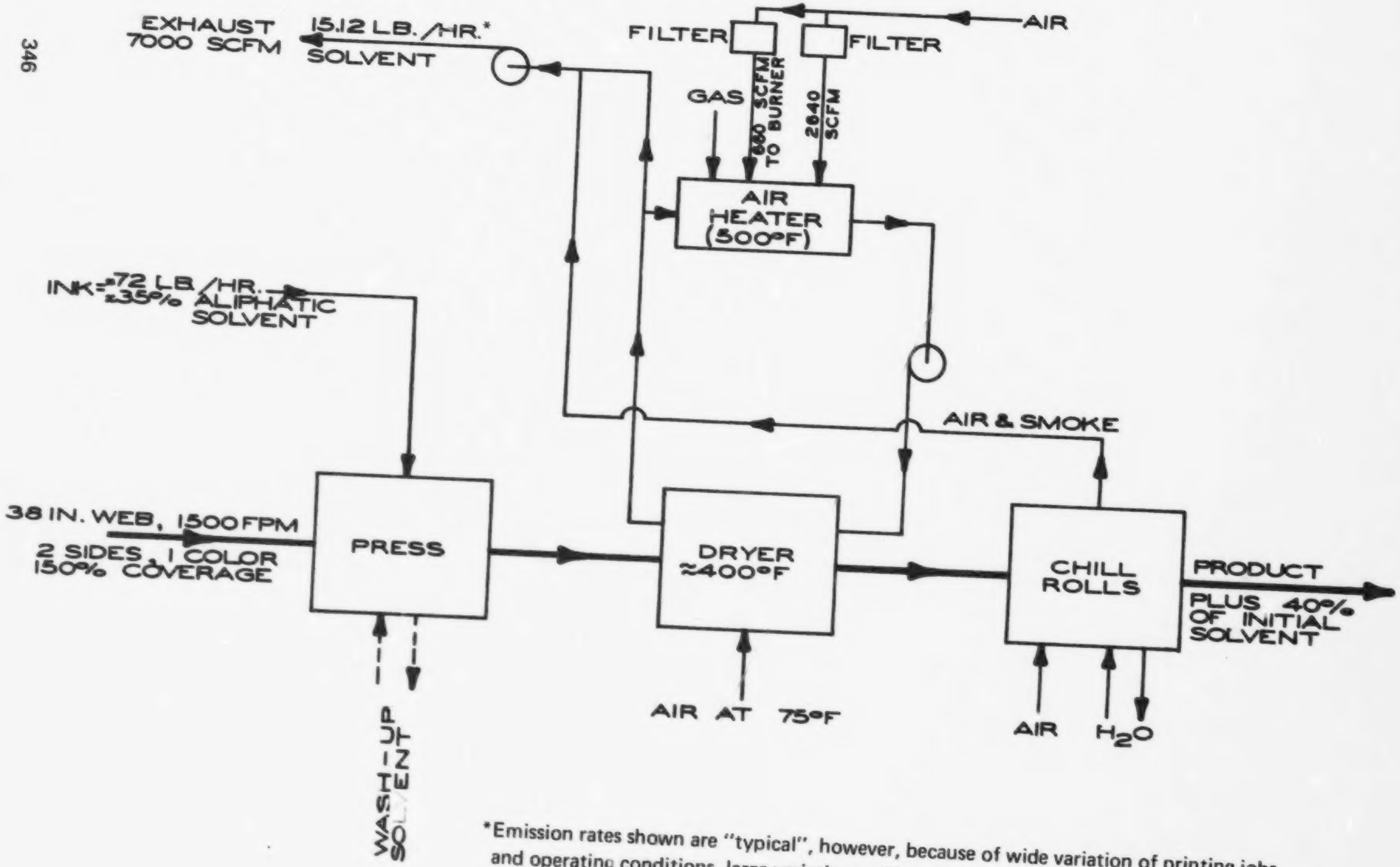


*Emission rates shown are "typical", however, because of wide variations of printing jobs and operating conditions, large variations will exist. The basis used for calculating emission rates is the following:

Web width	38 in.	
Number of sides printed	2	
Ink coverage	150%	
Ink mileage	$\frac{0.41\text{ lb ink}}{10^6\text{ in.}^2}$ coated,	$\frac{0.62\text{ lb ink}}{10^6\text{ in.}^2}$ non-coated
	Offset	
	$\frac{0.58\text{ lb ink}}{10^6\text{ in.}^2}$ coated,	$\frac{0.79\text{ lb ink}}{10^6\text{ in.}^2}$ non-coated
Solvent content of ink	35% – 60% removed during drying.	

FIGURE 87

WEB LETTERPRESS, PUBLICATION


EXHAUST
7000 SCFM
15.12 LB./HR.*
SOLVENT
FILTER
FILTER
AIR
GAS
660 SCFM
TO BURNER
2640
SCFM
AIR
HEATER
(500°F)
INK=72 LB./HR.
≈35% ALIPHATIC
SOLVENT
AIR & SMOKE
38 IN. WEB, 1500 FPM
2 SIDES, 1 COLOR
150% COVERAGE
PRESS
DRYER
≈400°F
CHILL
ROLLS
PRODUCT
PLUS 40%
OF INITIAL
SOLVENT
WASH-UP
SOLVENT
AIR AT 75°F
AIR
H2O

*Emission rates shown are "typical", however, because of wide variation of printing jobs and operating conditions, large variations will exist.

composition of the effluent gas also depends on the type of dryer employed. Emission rates as a function of press speed for a letterpress, heatset ink operation are shown in Figure 86 for the following operating conditions.

Web Width	38 in.
Number of Sides Printed	2
Ink Coverage	150%
Ink Mileage	0.58 lb ink/10^6 in.^{2}
% of Initial Solvent Removed in Dryer	60%

Web-Offset and Letterpress Newspaper Publication

Web-offset and letterpress newspaper printing operations employ oxidative drying inks which contain little, if any, solvent; therefore, the exhaust gases from these operations are not a source of hydrocarbon emissions. Usually the only materials emitted from these operations are ink mist and paper dust, which may be removed with filters as shown in Figure 88.

Rotogravure

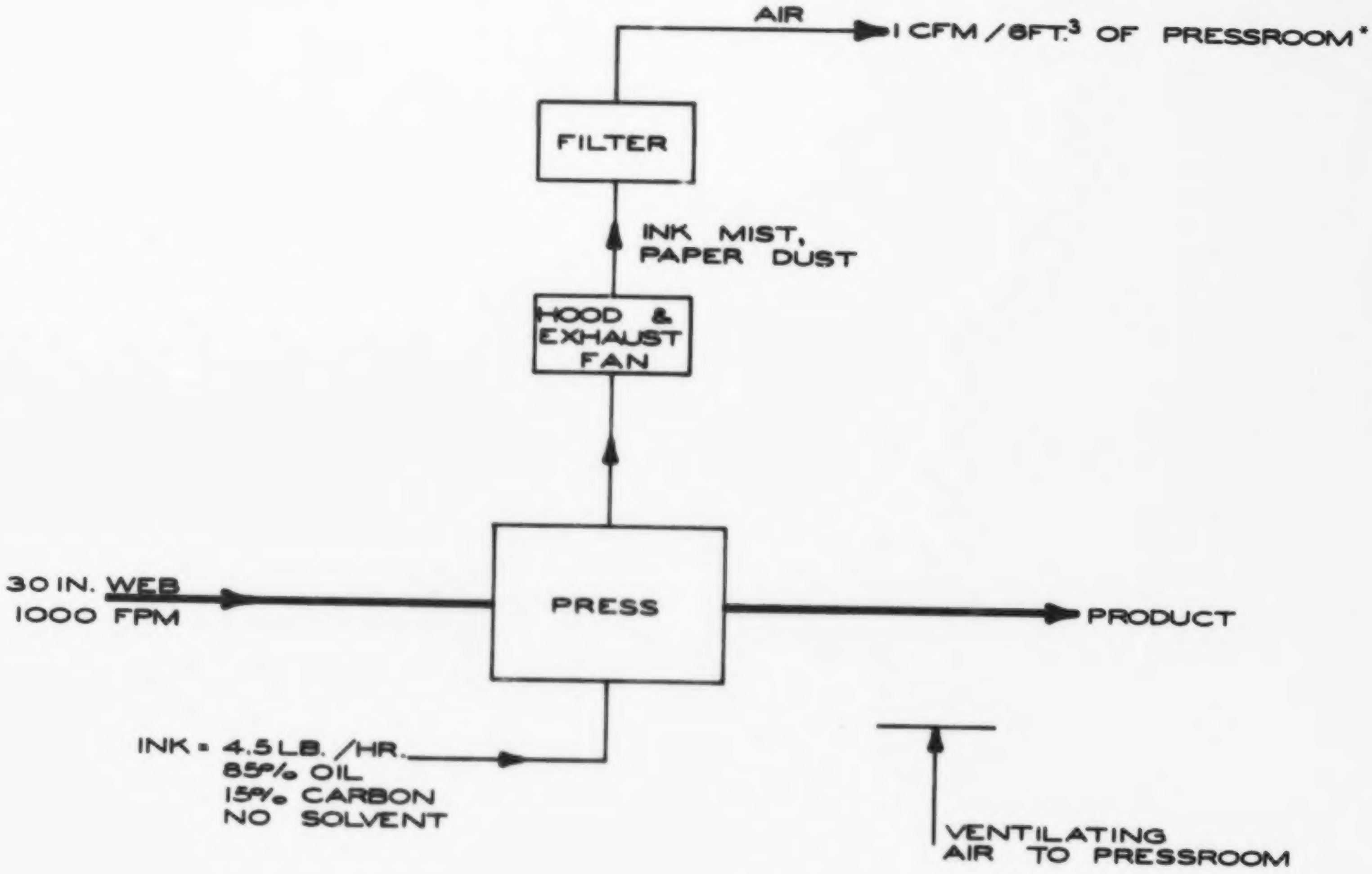
Rotogravure printing is similar to web-letterpress in that the web is printed on one side at a time and must be dried after each color is printed. In publication printing, the web is usually passed through four presses where four colors are applied to one side of the web. The web is then turned over and passed through four additional presses for the reverse side printing. Consequently, for four-color, two-sided printing, a total of eight presses must be employed, and each press will include a steam drum or hot air dryer where nearly all of the initial solvent is removed.

A solvent flow diagram for a single press operation is contained in Figure 89. For an integrated eight press operation, all of the dryers would be manifolded to one or two exhaust stacks; therefore, the exhaust rate shown in Figure 89 should be multiplied by four or eight for an actual operation.

As previously described, rotogravure inks are composed of approximately 65% highly volatile, aromatic solvent which is not subject to decomposition in the drying process. For most commercial operations, the solvent concentrations in the exhaust gases range between 25 and 40% of the lower explosive limit (LEL). Solvent emission rates as a function of press speed for a "typical" rotogravure printing operation running at the following conditions are contained in Figure 90.

FIGURE 88

WEB LETTERPRESS, NEWSPAPER

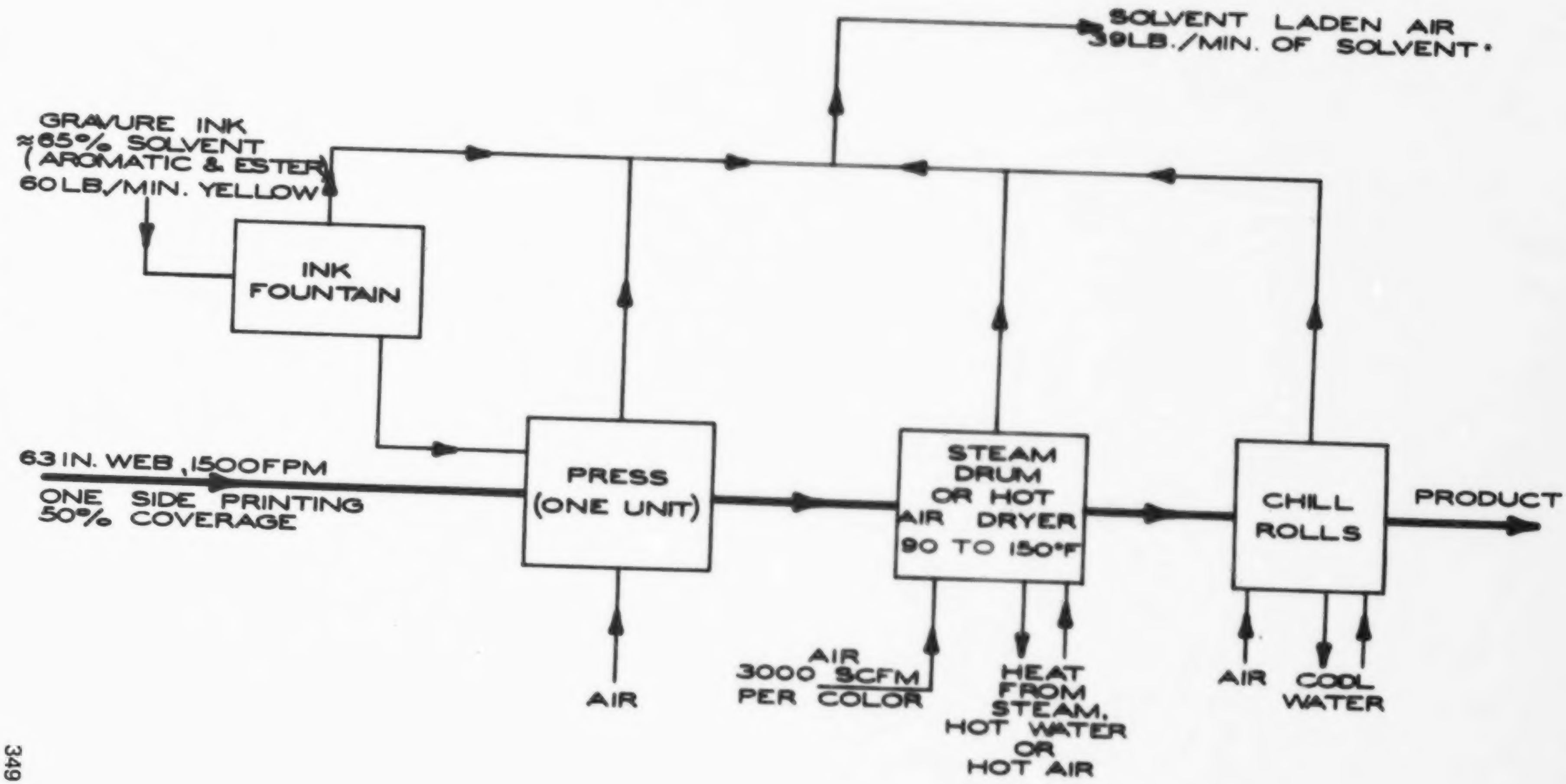

AIR
1 CFM / 6FT.³ OF PRESSROOM*
FILTER
INK MIST, PAPER DUST
HOOD & EXHAUST FAN
30 IN. WEB
1000 FPM
PRESS
PRODUCT
INK = 4.5 LB. /HR.
85% OIL
15% CARBON
NO SOLVENT
VENTILATING AIR TO PRESSROOM

---

*Emission rates shown are "typical", however, because of wide variation of printing jobs and operating conditions, large variations will exist.

FIGURE 89

ROTOGRAVURE PRINTING OPERATION


SOLVENT LADEN AIR
39LB./MIN. OF SOLVENT*
GRAVURE INK
≈65% SOLVENT
(AROMATIC & ESTER)
60LB./MIN. YELLOW
INK FOUNTAIN
63 IN. WEB, 1500FPM
ONE SIDE PRINTING
50% COVERAGE
PRESS (ONE UNIT)
STEAM DRUM OR HOT AIR DRYER 90 TO 150°F
CHILL ROLLS
PRODUCT
AIR
AIR 3000 SCFM PER COLOR
HEAT FROM STEAM, HOT WATER OR HOT AIR
AIR
COOL WATER

*Emission rates shown are "typical", however, because of wide variation of printing jobs and operating conditions, large variations will exist.

Web Width	63 in.
Number of Sides Printed	2
Number of Colors	4
Ink Coverage	125%
Ink Mileage	10.67 lb ink/10^6 in.^{2}
Solvent Content of Ink	65%

Metal Decorating

The metal decorating industry includes all operations involved in the printing of an image onto a sheet of metal. The printing image is usually applied to a coated sheet of metal with a lithographic press, sometimes followed by a final coating applied over the wet ink. The entire process involves three operations: application of the undercoating to the bare metal, application of the printing image to the dried coating, and application of the final overcoating to the wet image. The steps involved in a common three-stage metal decorating operation are illustrated in Figures 91 and 92.

The coating is usually a lacquer containing approximately 70% solvent which is applied with a roller type device (Figure 91). The sheets are then passed through a wicket oven where they are contacted with hot air at approximately 375°F. The quantity of solvent emitted from the oven is dependent on the thickness and solvent content of the lacquer coating; however, the ovens are usually operated from 10 to 25% of the lower explosive limit. If the metal must be coated on both sides, the sheets are coated and dried on one side, turned over, and the procedure is repeated.

The second and third steps (Figure 92) include the printing of the image onto the dry coated sheet of metal and the application of the overcoating. The coated sheets are printed with lithographic inks containing very little if any solvent. The wet inked sheets are then coated with a varnish containing approximately 50% solvent and dried in wicket ovens (≈320°F) which again are operated at 10 to 25% of the lower explosive limit.

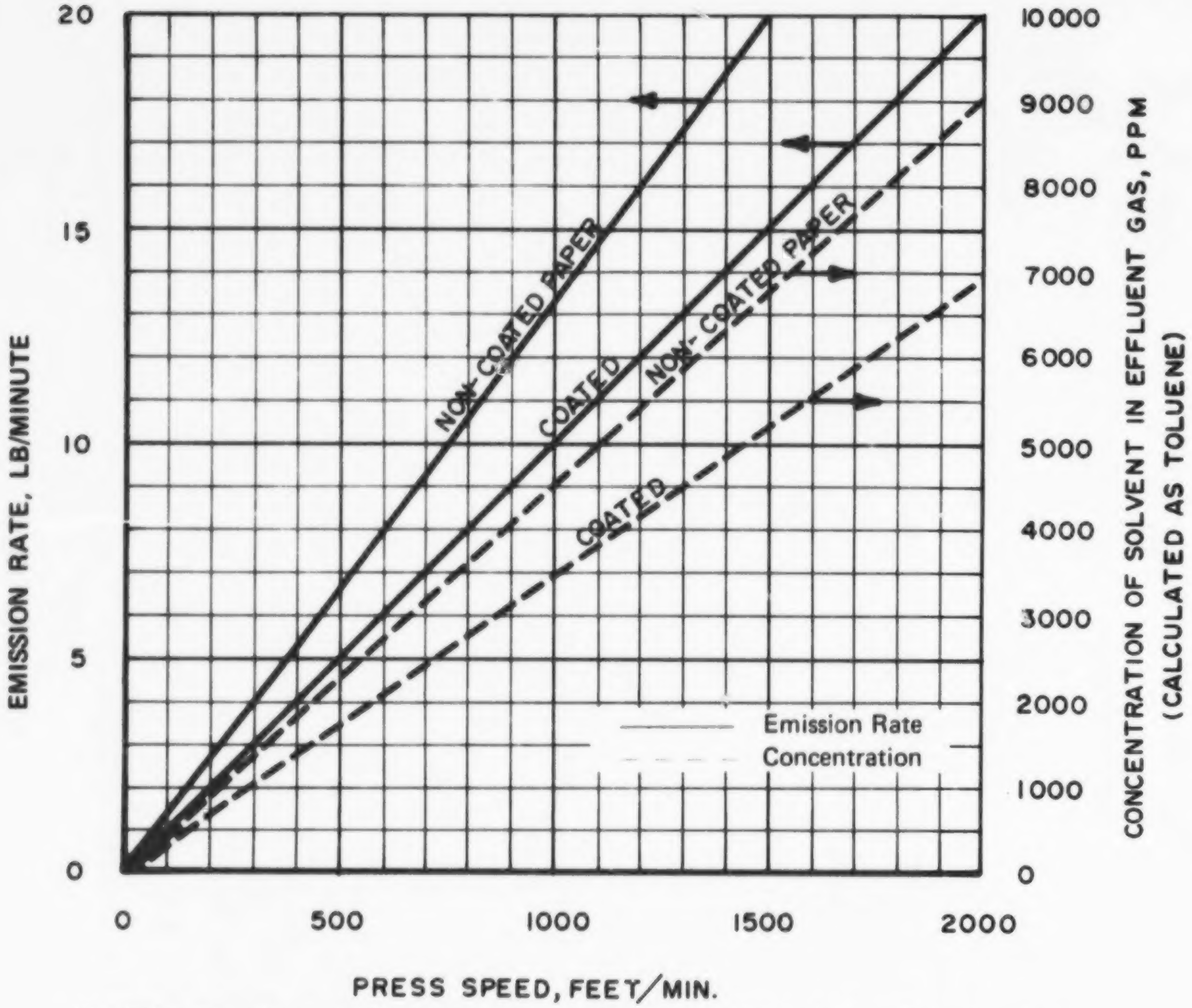
Since many metal decorating operations are for containers, three piece being older and more common than two piece, it should be mentioned that within some plants there are inside container spraying and baking emissions following seaming of the container body after the undercoating and printing. Frequently this operation is carried out by the packager at his location and since it is not directly involved in the printing operation, it is not considered in this study.

The newer two piece containers are not handled in sheet form nor does the body require seaming and the handling is much different although the emissions are similar.

The solvents employed in the varnish and lacquer coatings are usually composed of methyl isobutyl ketone (MIBK), xylol, and aliphatic solvent, all of which are removed in the wicket ovens.(8) The extent of solvent decomposition is a function of the variation of temperature due to a variation of the mixing efficiencies of the hot and cold gases in the oven.

FIGURE 90

EMISSION RATES* FROM A TYPICAL ROTOGRAVURE PRINTING OPERATION

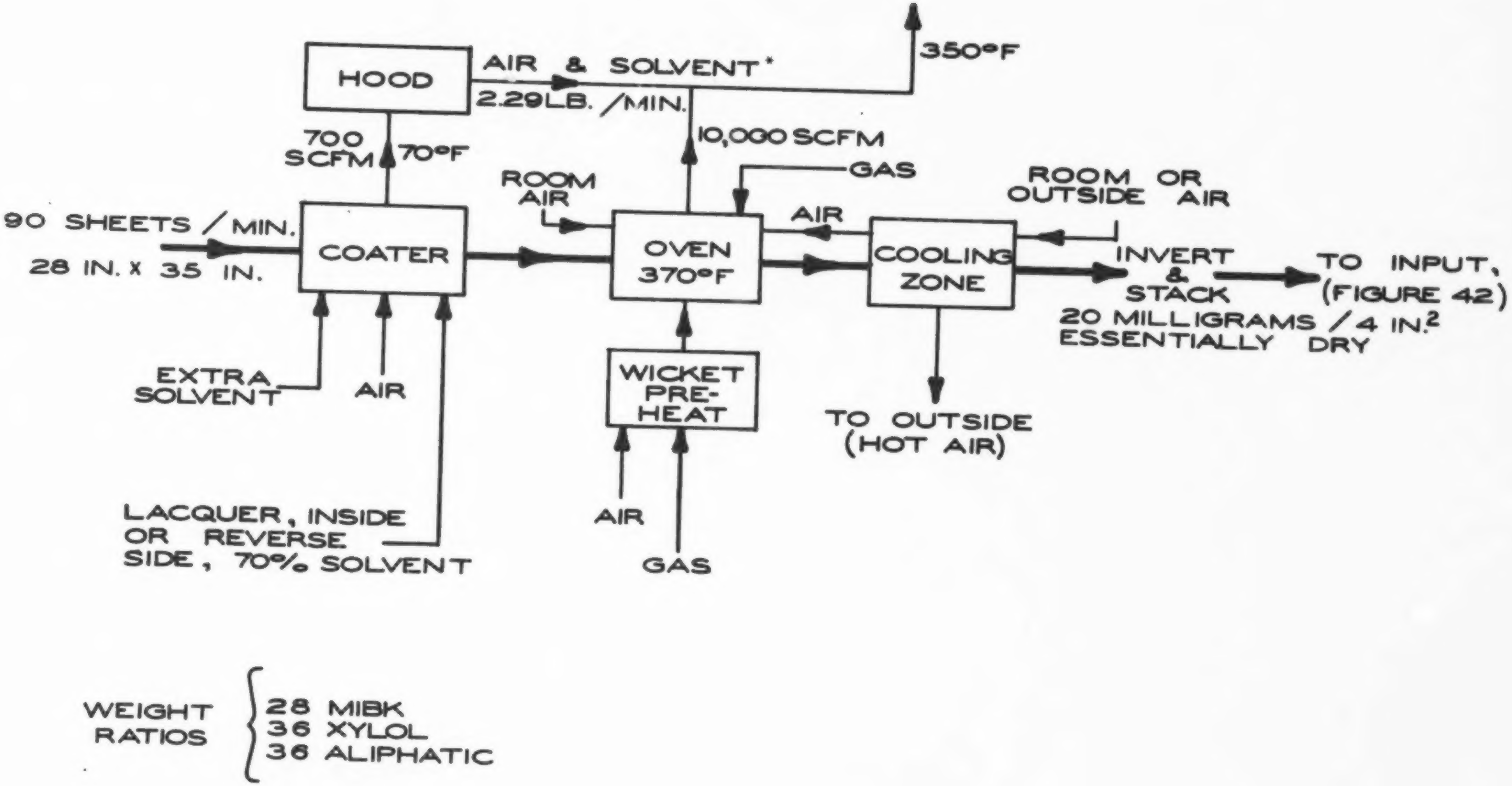


*Emission rates shown are "typical", however, because of wide variation of printing jobs and operating conditions, large variations will exist. The basis used for calculating emission rates is the following.

Web width	63"
Number of sides printed	2
Number of colors	4
Ink mileage – non-coated paper	8.00 lb ink/10^4 in.^{2}
Ink mileage – coated paper	10.67 lb ink/10^6 in.^{2}
Ink coverage	125%
Exhaust Rate	3000 SCFM/color
Solvent content of ink	65% – complete drying

FIGURE 91

METAL DECORATING
COATING OPERATION

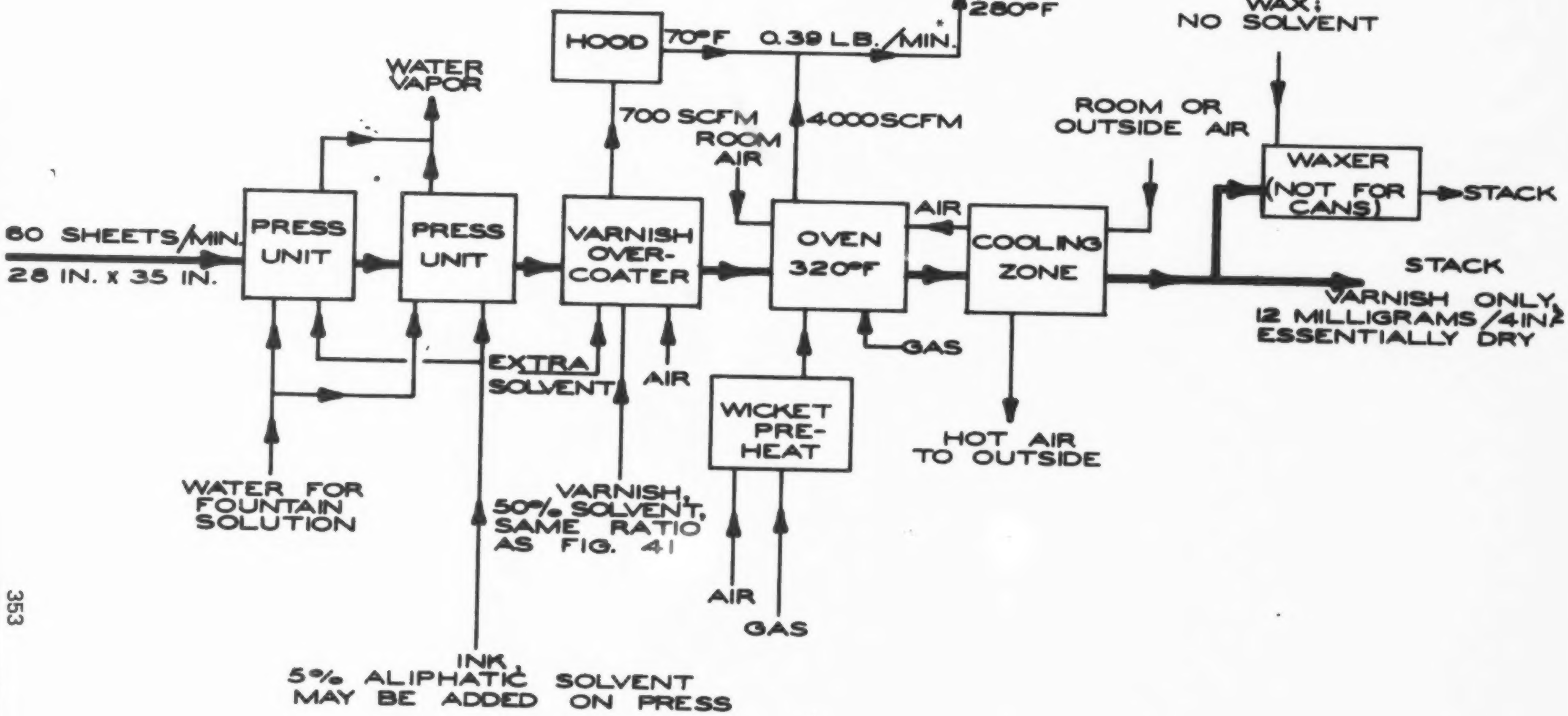

HOOD
AIR & SOLVENT*
2.29LB. /MIN.
350°F
700 SCFM
70°F
10,000 SCFM
GAS
ROOM OR OUTSIDE AIR
ROOM AIR
AIR
90 SHEETS / MIN.
28 IN. x 35 IN.
COATER
OVEN 370°F
COOLING ZONE
INVERT & STACK
TO INPUT, (FIGURE 42)
20 MILLIGRAMS / 4 IN.²
ESSENTIALLY DRY
EXTRA SOLVENT
AIR
WICKET PRE-HEAT
TO OUTSIDE (HOT AIR)
LACQUER, INSIDE OR REVERSE SIDE, 70% SOLVENT
AIR
GAS
WEIGHT RATIOS
28 MIBK
36 XYLOL
36 ALIPHATIC

---

*Emission rates shown are "typical", however, because of wide variation of printing jobs and operating conditions, large variations will exist.

FIGURE 92

METAL DECORATING
PRINTING AND VARNISH OVERCOATING



60 SHEETS/MIN.
28 IN. x 35 IN.
PRESS UNIT
PRESS UNIT
WATER VAPOR
WATER FOR FOUNTAIN SOLUTION
INK, 5% ALIPHATIC SOLVENT MAY BE ADDED ON PRESS
EXTRA SOLVENT
VARNISH OVER-COATER
VARNISH, 50% SOLVENT, SAME RATIO AS FIG. 41
AIR
HOOD
700 SCFM
70°F
0.39 LB./MIN.*
280°F
4000 SCFM
ROOM AIR
OVEN 320°F
GAS
WICKET PRE-HEAT
AIR
GAS
AIR
COOLING ZONE
ROOM OR OUTSIDE AIR
HOT AIR TO OUTSIDE
WAX; NO SOLVENT
WAXER (NOT FOR CANS)
STACK
STACK
VARNISH ONLY, 12 MILLIGRAMS/4IN2 ESSENTIALLY DRY

*Emission rates shown are "typical", however, because of wide variation of printing jobs and operating conditions, large variations will exist.

ABATEMENT SYSTEMS

In the printing and metal decorating industries, gaseous emissions of hydrocarbons are potential sources of photochemical smog and control of these emissions can be accomplished. Possible control techniques applicable to the graphic arts industry may be classified into the following categories:

1. Process modification
2. Ink modification
3. Conventional air pollution control equipment

Process Modification

Since the major portion of hydrocarbon emissions from a printing operation is produced during the drying process, the most logical area for modification is the dryer. The type of dryer employed in a particular operation depends on the substrate, ink composition, press speed, and type of printing process and may consist of flame and high velocity air, heated air, or steam. Several novel methods of drying are currently being developed which may reduce or completely eliminate hydrocarbon emissions: microwave drying, ultraviolet drying, electron beam drying, and infrared drying.

In microwave drying, the temperature of the ink is increased by the application of electromagnetic energy (alternating electric field). Since fuel is not directly consumed in this type of operation, the oven exhaust will not contain combustion products; however, solvent vapors will still be emitted if conventional inks are used.(9)

In infrared drying, the ink is dried by a free radical polymerization mechanism which requires a non-volatile, monomer-based ink. Since the ink will not contain a volatile solvent, hydrocarbon emissions will be eliminated. This type of drying is still in the early experimental stages.

Electron beam drying requires inks composed of monomers or prepolymers which solidify by electron-induced polymerization. This procedure is currently being employed in a small number of commercial operations; however, equipment costs are high and since X-rays are produced when the electrons strike the target, shielding of the entire unit is required.

In ultraviolet drying, lamps operating at 2400 to 3600 angstroms activate monomer-based inks resulting in rapid polymerization. It is anticipated that ultraviolet drying will be employed in 80% of the offset printing operations and 70% of the letterpress printing operations within five years. Although hydrocarbon emissions are eliminated in this type of drying, the system does have some inherent disadvantages: the monomer-based inks are more expensive than conventional inks, ozone is produced in the process and must be vented, and the inks are not easily removed during paper reclamation.

Ink Modification

A great deal of research is currently underway to develop non-pollutant inks. The majority of the research is aimed at the development of aqueous and solventless inks.

Aqueous inks are already being used in some flexographic operations. One major disadvantage of the aqueous system is that the relatively high latent heat of water limits the press speeds when conventional dryers are employed; however, the application of microwave drying to aqueous systems has actually enabled press speeds to increase.(10)

Solventless inks are currently being marketed for web-offset and web-letterpress operations. In contrast to solvent base inks which are dried by evaporation, solventless inks are dried by a thermally-induced polymerization reaction which appreciably reduces or completely eliminates hydrocarbon emissions. The ink manufacturers claim that these inks can be adapted to present equipment without modification and, since lower oven temperatures are required, press speeds may be increased. Some problems have apparently been encountered in areas of gloss, sharpness of printing, and quality, and the ink manufacturers are actively pursuing solutions to these problems.

Conventional Air Pollution Control Equipment

Exhaust gases from printing and metal decorating plants may be treated with conventional pollution control equipment to eliminate the organic emissions. Three conventional abatement systems are applicable to the graphic arts industry, and the advantages and disadvantages of each will be discussed: thermal and catalytic combustion and solvent recovery (adsorption).

Thermal combustion involves burning the effluent gas directly in a gas or oil fired flame. The combustible components of the exhaust stream are heated to their auto-ignition temperature in the presence of sufficient oxygen to complete the following reaction:

$$C_NH_M + (N + M/2)\ O_2 \rightarrow N\ CO_2 + (M/2)\ H_2O + \text{Heat}$$

Since chemical reactions do not proceed instantaneously, sufficient residence time must be incorporated into the system to complete the reaction. The amount of time required increases as the reaction temperature decreases. If the combustion reaction is not carried to completion, intermediate combustion products are produced which may be more offensive than the original contaminants. Examples of incomplete oxidation are illustrated in Figure 93.

A schematic flow diagram of a thermal combustion system is shown in Figure 94. The oven exhaust gases are preheated, by some means, to 600 to 900° F depending on the composition of the gas. The hot gases are then incinerated at temperatures of 1200 to 1600° F. The exact amount of temperature rise across the incinerator depends on the heat of combustion and quantity of hydrocarbon vapors in the gas. Fuel consumption is dependent on the amount of heat exchange employed in the process and the operating temperature. Thermal incinerators are capable of operating continuously at efficiencies of 90 to 99%.

Catalytic combustion is a flameless oxidation reaction which is accelerated by means of a catalyst. The catalyst usually consists of a platinum group metal deposited on a ceramic base or metal ribbon. Since the catalysts are subject to thermal deactivation and poisoning, catalytic combustion usually requires closer observation than thermal combustion. Well known poisons for platinum group catalysts are lead, zinc, and silicones which, when present in small quantities, will severely, if not completely, deactivate the catalyst. Operation of catalytic incinerators above 1100° F will result in shorter catalyst life, higher maintenance, and consequently, higher operating costs.

A schematic flow diagram of a catalytic incinerator installation is contained in Figure 95. The flow diagram is very similar to a thermal incinerator except for the lower reaction temperature. Efficiencies usually range between 85 and 95% depending on the application.(2)

The removal of hydrocarbon vapors from a gas stream by adsorption onto a solid is becoming increasingly popular, primarily because the hydrocarbons may be recovered and reused. In a typical solvent recovery operation (Figure 96), the vapor-laden gas is passed through a bed of activated carbon until the

FIGURE 93

PRODUCTS PRODUCED AT VARIOUS OXIDATION STAGES DURING COMBUSTION

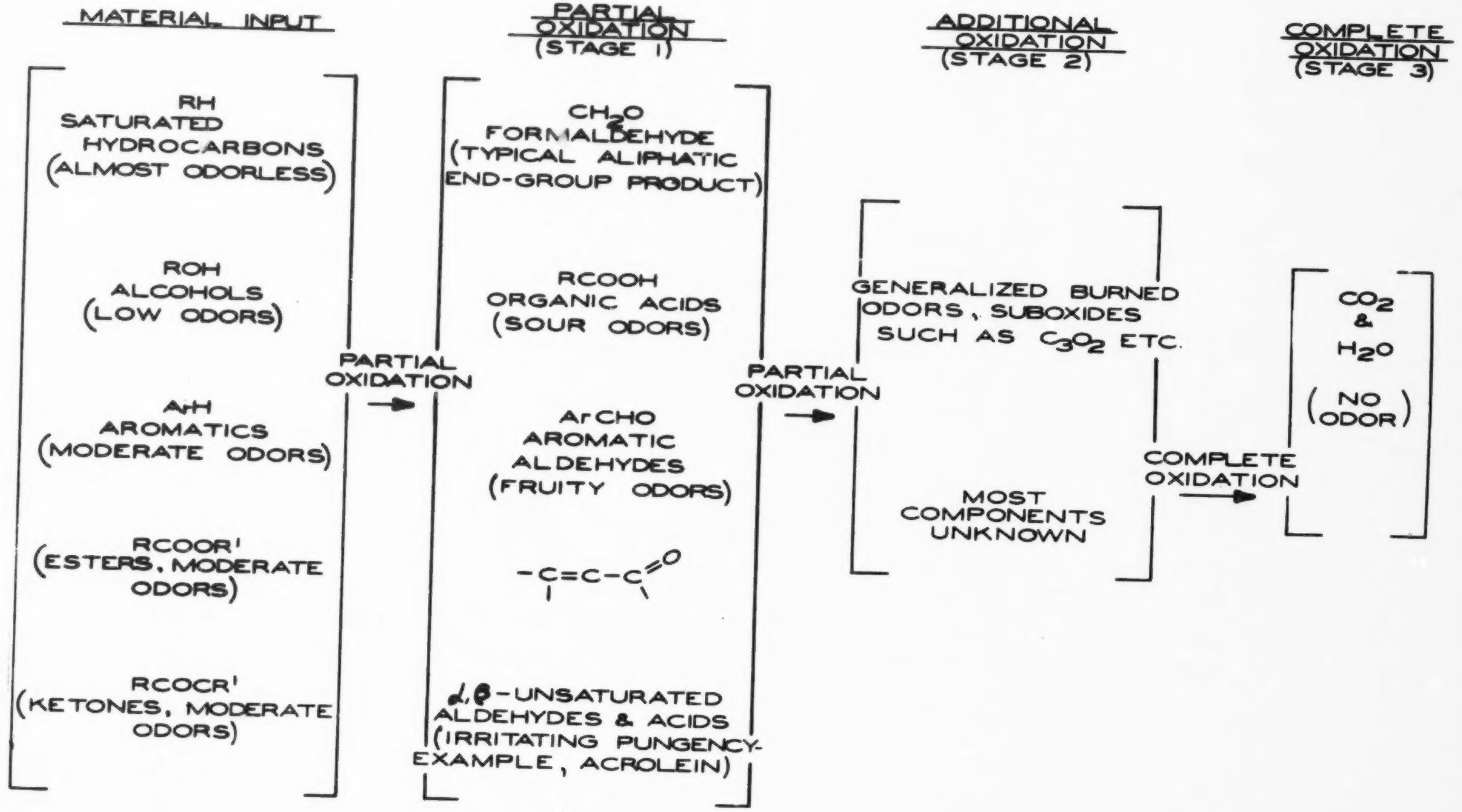

MATERIAL INPUT
PARTIAL OXIDATION (STAGE 1)
ADDITIONAL OXIDATION (STAGE 2)
COMPLETE OXIDATION (STAGE 3)
RH SATURATED HYDROCARBONS (ALMOST ODORLESS)
ROH ALCOHOLS (LOW ODORS)
ArH AROMATICS (MODERATE ODORS)
RCOOR' (ESTERS, MODERATE ODORS)
RCOCR' (KETONES, MODERATE ODORS)
PARTIAL OXIDATION
CH2O FORMALDEHYDE (TYPICAL ALIPHATIC END-GROUP PRODUCT)
RCOOH ORGANIC ACIDS (SOUR ODORS)
ArCHO AROMATIC ALDEHYDES (FRUITY ODORS)
-C=C-C=O
α,β-UNSATURATED ALDEHYDES & ACIDS (IRRITATING PUNGENCY-EXAMPLE, ACROLEIN)
PARTIAL OXIDATION
GENERALIZED BURNED ODORS, SUBOXIDES SUCH AS C3O2 ETC.
MOST COMPONENTS UNKNOWN
COMPLETE OXIDATION
CO2 & H2O (NO ODOR)

FIGURE 94

FLOW DIAGRAM FOR THERMAL COMBUSTION INCLUDING POSSIBILITIES FOR HEAT RECOVERY

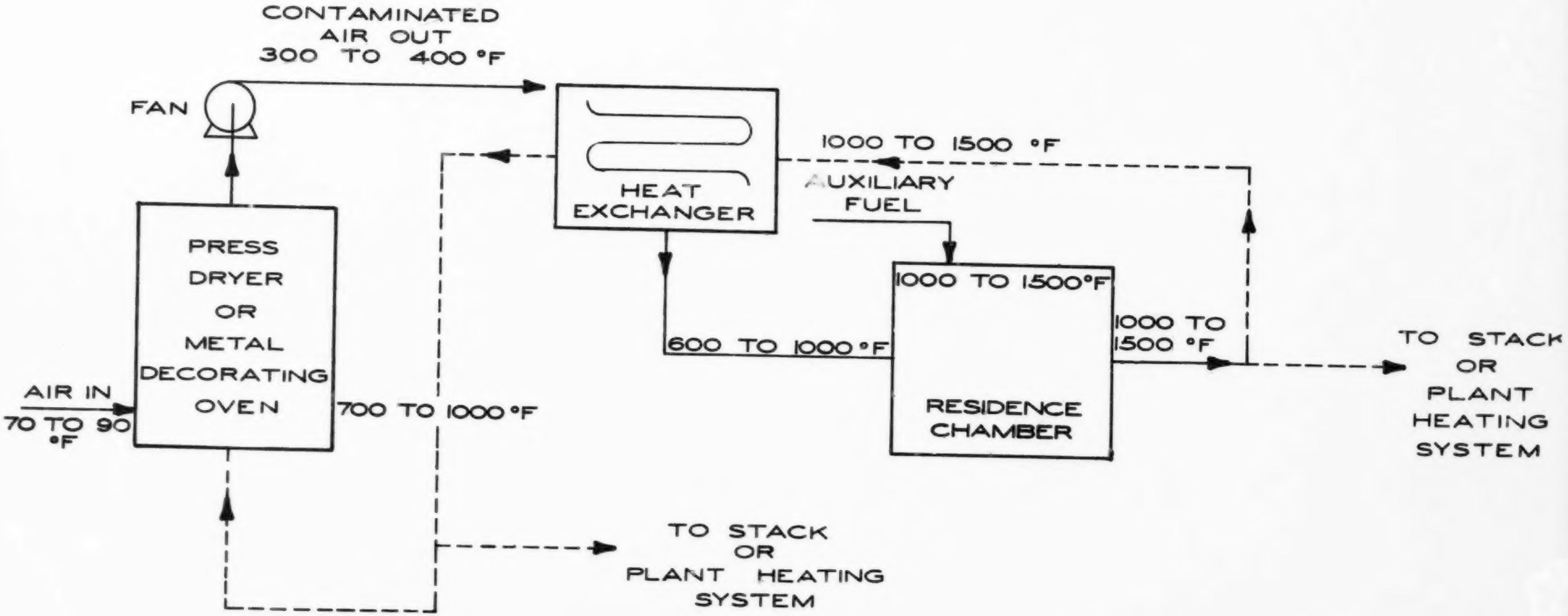

CONTAMINATED
AIR OUT
300 TO 400 °F
FAN
PRESS
DRYER
OR
METAL
DECORATING
OVEN
AIR IN
70 TO 90 °F
700 TO 1000 °F
HEAT
EXCHANGER
1000 TO 1500 °F
AUXILIARY
FUEL
1000 TO 1500°F
600 TO 1000 °F
RESIDENCE
CHAMBER
1000 TO
1500 °F
TO STACK
OR
PLANT
HEATING
SYSTEM
TO STACK
OR
PLANT HEATING
SYSTEM

FIGURE 95

FLOW DIAGRAM FOR CATALYTIC COMBUSTION INCLUDING POSSIBILITIES FOR HEAT RECOVERY

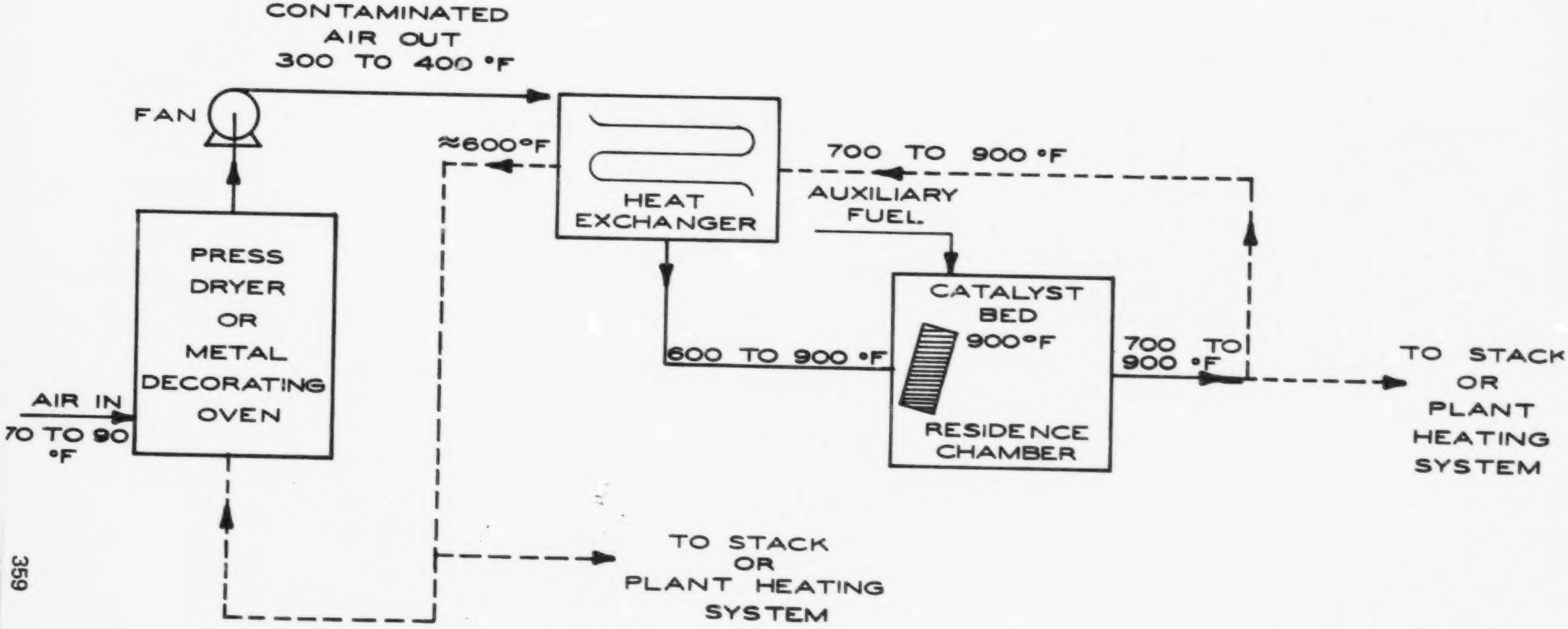
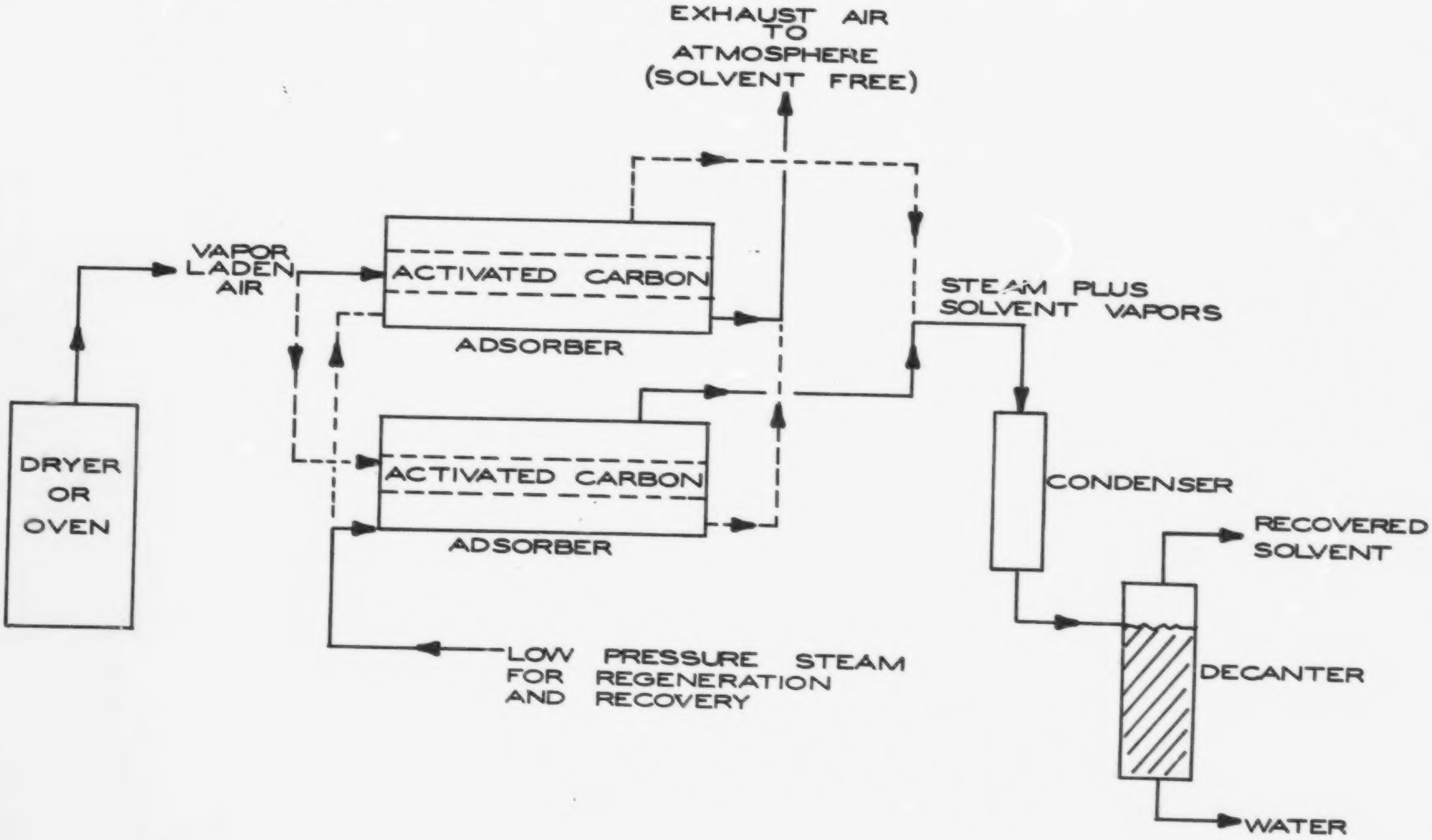

CONTAMINATED
AIR OUT
300 TO 400 °F
FAN
≈600°F
HEAT
EXCHANGER
700 TO 900 °F
AUXILIARY
FUEL
PRESS
DRYER
OR
METAL
DECORATING
OVEN
AIR IN
70 TO 90
°F
CATALYST
BED
900°F
600 TO 900 °F
RESIDENCE
CHAMBER
700 TO
900 °F
TO STACK
OR
PLANT
HEATING
SYSTEM
TO STACK
OR
PLANT HEATING
SYSTEM

FIGURE 96

FLOW DIAGRAM OF ADSORPTION PROCESS

ADSORPTION (SOLVENT-RECOVERY SYSTEM)


EXHAUST AIR
TO
ATMOSPHERE
(SOLVENT FREE)
VAPOR
LADEN
AIR
ACTIVATED CARBON
ADSORBER
ACTIVATED CARBON
ADSORBER
DRYER
OR
OVEN
STEAM PLUS
SOLVENT VAPORS
CONDENSER
RECOVERED
SOLVENT
DECANTER
WATER
LOW PRESSURE STEAM
FOR REGENERATION
AND RECOVERY

SWITCHING OF AIR FLOWS WOULD BE NECESSARY TO ACHIEVE ADSORPTION - REGENERATION CYCLE adsorption capacity of the bed has been reached. The gas stream is then diverted to an alternate bed while the original bed is regenerated with steam or hot air. If the hydrocarbon solvent is not miscible with water, it can be recovered by simple decantation; otherwise, recovery of the hydrocarbon will require distillation.

Adsorption is governed by the laws of chemical equilibrium and diffusion and, therefore, the following operation variables will influence the capacity of a carbon bed for an organic solvent:

Partial Pressure –	the adsorption capacity of a bed increases as the partial pressure of the vapor in the gas stream increases.
Temperature –	the capacity of the bed increases as the temperature decreases.
Molecular Size –	the bed capacity decreases as the molecular size increases.
Water Vapor –	water vapor decreases the capacity of a bed.

Performance will also be influenced by the presence of and the nature of solid and liquid particulate matter in the gas stream being treated. In most designs, filters precede the carbon beds in order to remove suspended particulate matter.

Generally, in a well-designed adsorption unit, the activated carbon will adsorb approximately 15% of its own weight of solvent before regeneration is required.

SPECIFICATIONS AND COSTS

Abatement specifications for three sets of graphic arts applications were written: web offset printing, metal decorating, and gravure printing. In the case of web offset printing, both thermal and catalytic combustion systems without heat exchange were specified. Heat exchange was not included in the specifications because the ventilation rates from the model printing presses studied were too low. Cost data and equipment specifications for the web offset applications are presented in Tables 125 to 132 and in Figures 97 to 103.

Thermal and catalytic combustion systems were also specified for the metal decorating application. Since the gas flows were higher for the metal decorating cases, however, systems both with and without heat exchange were specified. Cost data and equipment specifications are presented in Tables 133 to 148 and in Figures 103 to 114.

Gravure printing applications require that large volumes of ventilation air be treated relative to metal decorating or web offset printing applications. Specifications were written for two types of abatement systems: thermal combustion systems and activated carbon adsorption systems. The thermal combustion system specifications were written to apply to one or two presses in the printing plant. This is in keeping with the industry practice of using one incinerator for every one or two press vents. The flow from each gravure press vent is large enough to clearly justify heat exchange in the combustion system and the specifications have been written accordingly. Cost data and equipment specifications are presented in Tables 149 to 152 and in Figures 115 to 117.

The carbon adsorption system specifications were written to apply to all of the presses in a printing plant. The model plants studied in this contract contained four and twelve roughly equivalent presses. Cost data and equipment specifications are presented in Tables 153 to 157 and in Figures 118 to 120. The carbon adsorption system differs from the other abatement systems studied in that it recovers the hydrocarbon emissions in a form reusable as solvent in the gravure printing process. The value of the recovered solvent is significant relative to the total operating costs of the adsorption system and, in the cases studied here, offers an attractive payout on the capital investment. The recovered solvent credit, valued at $0.04/lb is shown with the operating cost data in Table 156 and Figure 119.

Preceding page blank

TABLE 125

THERMAL INCINERATOR PROCESS DESCRIPTION FOR WEB-OFFSET PRINTING SPECIFICATION

This specification describes the requirements of a thermal combustion system for the abatement of hydrocarbon emissions from a web-offset printing facility. A schematic flow diagram of the desired combustion system is included in Figure 97 while processing conditions and specifications for small and large facilities are tabulated in Table 126.

The incinerator will be gas fired using natural gas available at a pressure of 1.0 psig and having a specific gravity of 0.6 and an upper heating value of 1,040 Btu/SCF. The exhaust gas from the printing press will contain sufficient oxygen (greather than 16%) to allow firing of the burner without the addition of a combustion air system.

A fan equipped with a V-belt drive will be required at the incinerator inlet. The fan will have the capacity to overcome the pressure drop of the ductwork and the system. The system's ductwork will be sized for a maximum ΔP *of 2 in. w.c. (hot).*

The incinerator will be supplied with suitable control panels and all equipment will be designed for outdoor operation and to meet Factory Insurance Association's standards.

The cost estimate shall include the following items:

Incinerator
Burner
10 ft stack
Controls
Control panel
Structural steel
Fuel gas piping
Electrical
Ductwork
Insulation
Fan
Fan motor
Two day start-up service

All items with exception of the incinerator and burner shall be considered auxiliaries.

The unit will be designed for the maximum hydrocarbon concentration, while operating costs shall be estimated at the average hydrocarbon concentration.

The incinerator will be located on the roof of the printing facility and no modification of the building structure is required. All utilities are available within 30 ft of the control cabinet, motor, and burner.

FIGURE 97

SCHEMATIC FLOW DIAGRAM OF
THERMAL AND CATALYTIC
COMBUSTION WITHOUT
HEAT EXCHANGER FOR
WEB-OFF SET AND METAL DECORATING

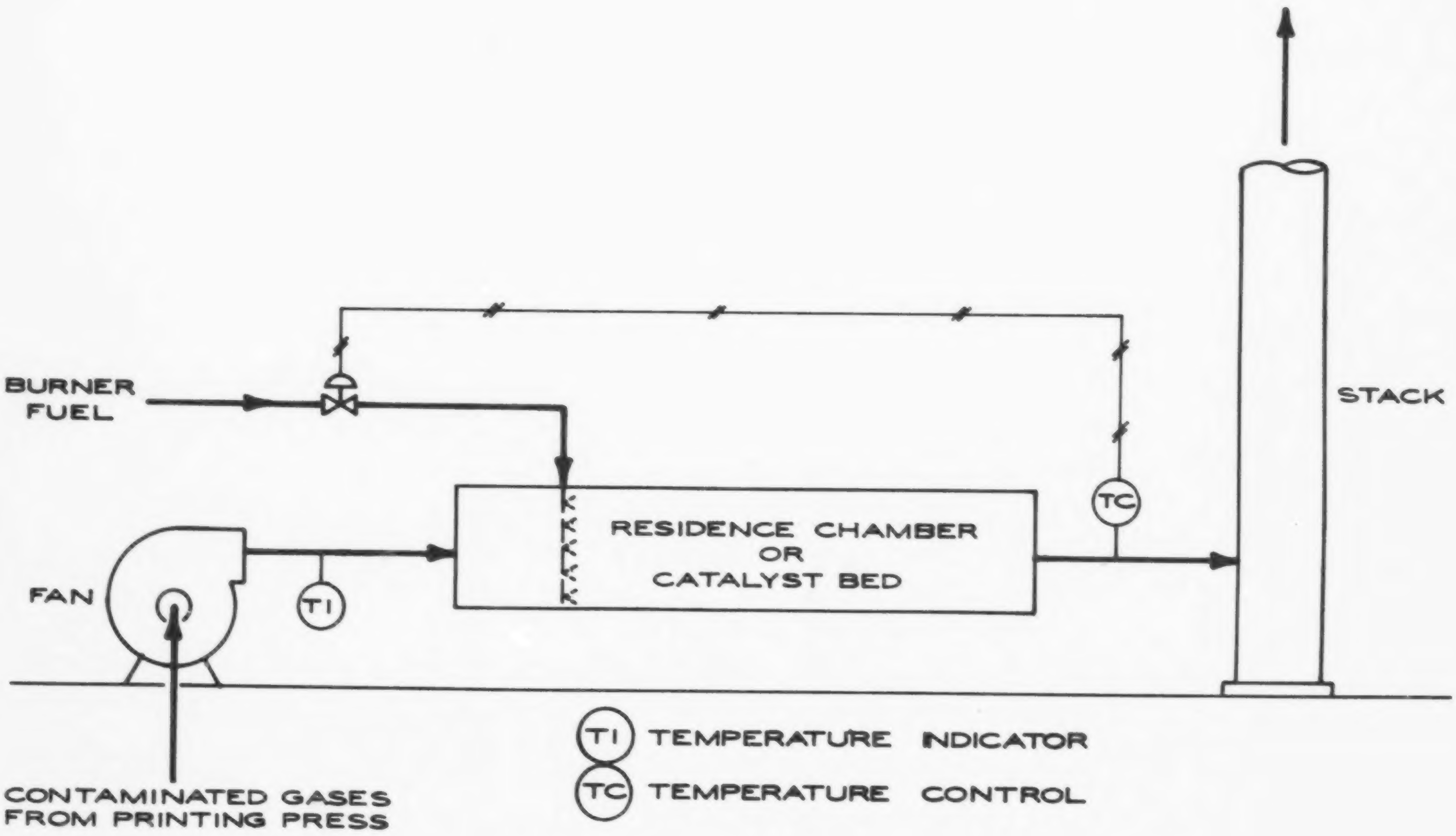

BURNER
FUEL
FAN
TI
RESIDENCE CHAMBER
OR
CATALYST BED
TC
STACK
CONTAMINATED GASES
FROM PRINTING PRESS
TI TEMPERATURE INDICATOR
TC TEMPERATURE CONTROL

TABLE 126

THERMAL INCINERATOR OPERATING CONDITIONS FOR WEB-OFFSET PRINTING SPECIFICATION (WITHOUT HEAT EXCHANGE)

Plant Size	*Small*	*Large*
Process Conditions		
Effluent Gas Temperature, °F	*350*	*350*
Effluent Gas Rate, SCFM	*2,000*	*7,000*
Effluent Gas Rate, ACFM	*3,060*	*10,700*
Hcbn Concentration, ppm(1)	*500-2,000*	*500-2,000*
Hcbn Emission Rate, lb/hr		
Maximum	*10.0*	*35.4*
Average	*6.3*	*22.1*
Heat of Combustion of Hcbn, Btu/lb(2)	*19,500*	*19,500*
Heating Value of Gas, Btu/SCF		
Maximum	*1.5*	*1.5*
Average	*0.9*	*0.9*
Incinerator Specifications		
Inlet Temperature, °F	*350*	*350*
Outlet Temperature, °F	*1,500*	*1,500*
Residence Time @ Temperature, sec	*0.6*	*0.6*
*Burner Duty, MMBtu/hr**	*3.35*	*10.52*

(1) Measured as methane equivalents
(2) Assumed as n-hexane

**To be filled in by supplier*

Preceding page blank

TABLE 127

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR WEB-OFFSET PRINTING (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			3,060	10,700
°F			350	350
SCFM			2,000	7,000
Moisture Content, Vol. %			-	-
Effluent Contaminant Loading				
ppm			500-2,000	500-2,000
lb/hr avg.			6.3	22.1
Btu/scf			0.9	0.9
Cleaned Gas Flow				
ACFM			7,550	26,480
°F			1,500	1,500
SCFM			2,039	7,136
Moisture Content, Vol. %			5.35	5.35
Cleaned Gas Contaminant Loading				
ppm			5-20	5-20
lb/hr			<0.3	<1.1
Cleaning Efficiency, %			>95	>95
(1) Gas Cleaning Device Cost			16,075	21,025
(2) Auxiliaries Cost				
(a) Fan(s)			1,870	3,420
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			1,950	2,450
(b) Foundations & Support			1,200	1,625
Ductwork			2,150	2,850
Stack			650	850
Electrical			625	800
Piping			300	350
Insulation				
Painting				
Supervision			600	600
Startup			450	450
Performance Test			1,750	1,750
Other			2,450	2,688
(4) Total Cost			30,070	38,858

FIGURE 98

CAPITAL COSTS FOR THERMAL INCINERATORS
FOR WEB-OFFSET PRINTING
(WITHOUT HEAT EXCHANGE)

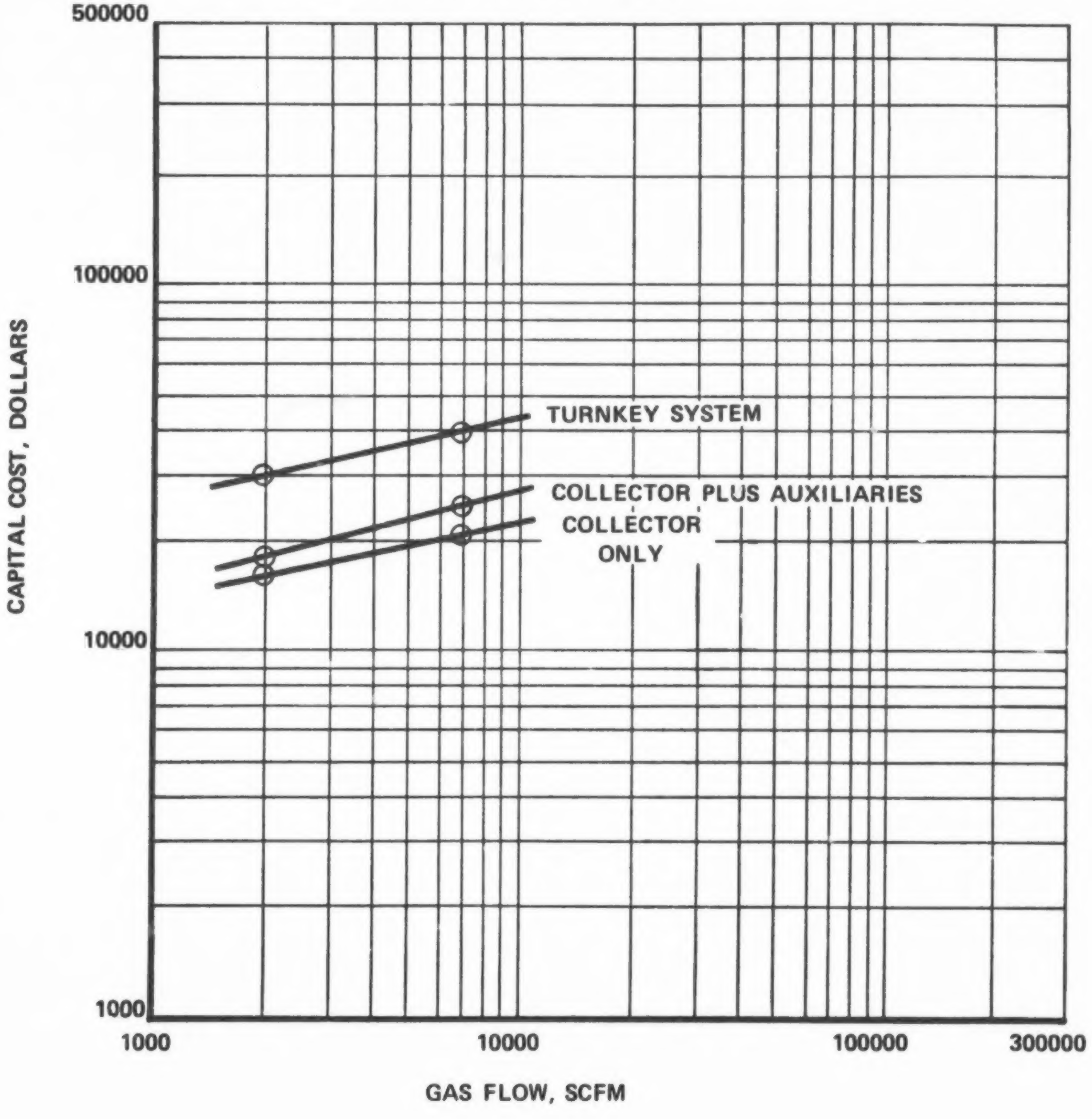


TABLE 128

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR THERMAL INCINERATORS FOR WEB-OFFSET PRINTING (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				4,000	4,000
Operating Labor (if any)					
Operator	$6/hr			1,500	1,500
Supervisor					
Total Operating Labor				1,500	1,500
Maintenance					
Labor	$6/hr			336	336
Materials				135	160
Total Maintenance				471	496
Replacement Parts				125	125
Total Replacement Parts				125	125
Utilities					
Electric Power	$.011/kw-hr			222	813
Fuel	$.80/MMBtu			10,026	33,196
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				10,248	34,009
Total Direct Cost				12,344	36,130
Annualized Capital Charges				3,007	3,886
Total Annual Cost				15,351	40,016

FIGURE 99

ANNUAL COSTS FOR THERMAL INCINERATORS FOR WEB-OFFSET PRINTING (WITHOUT HEAT EXCHANGE)

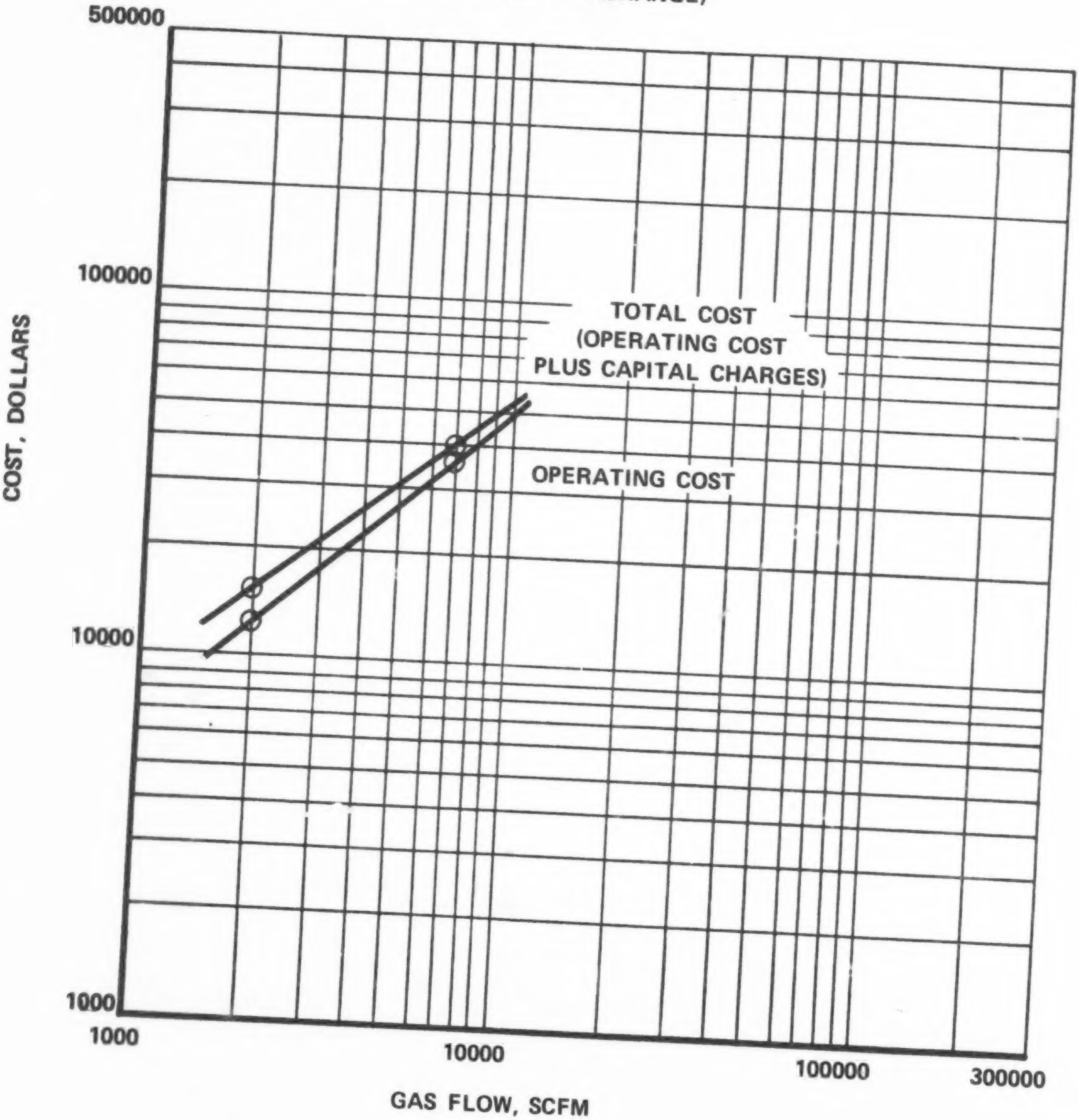


TABLE 129

CATALYTIC INCINERATOR PROCESS DESCRIPTION FOR WEB-OFFSET PRINTING SPECIFICATION

This specification describes the requirements of a thermal combustion system for the abatement of hydrocarbon emissions from a web-offset printing facility. A schematic flow diagram of the desired combustion system is included in Figure 100 while processing conditions and specifications for small and large facilities are tabulated in Table 130.

The incinerator will be gas fired using natural gas available at a pressure of 1.0 psig and having a specific gravity of 0.6 and an upper heating value of 1,040 Btu/SCF. The exhaust gas from the printing press will contain sufficient oxygen (greather than 16%) to allow firing of the burner without the addition of a combustion air system.

A fan equipped with a V-belt drive will be required at the incinerator inlet. The fan will have the capacity to overcome the pressure drop of the ductwork and the system. The system's ductwork will be sized for a maximum Δ P of 2 in. w.c. (hot).

The incinerator will be supplied with suitable control panels, and all equipment will be designed for outdoor operation and to meet Factory Insurance Association's standards.

The cost estimate shall include the following items:

Incinerator
Burner
10 ft stack
Controls
Control panel
Structural steel
Fuel gas piping
Electrical
Ductwork
Insulation
Fan
Fan motor
Two day start-up service

All items with exception of the incinerator and burner shall be considered auxiliaries.

The unit will be designed for the maximum hydrocarbon concentration, while operating costs shall be estimated at the average hydrocarbon concentration.

The incinerator will be located on the roof of the printing facility and no modification of the building structure is required. All utilities are available within 30 ft of the control cabinet, motor, and burner.

FIGURE 100

SCHEMATIC FLOW DIAGRAM OF THERMAL AND CATALYTIC COMBUSTION WITHOUT HEAT EXCHANGE FOR WEB-OFF SET AND METAL DECORATING

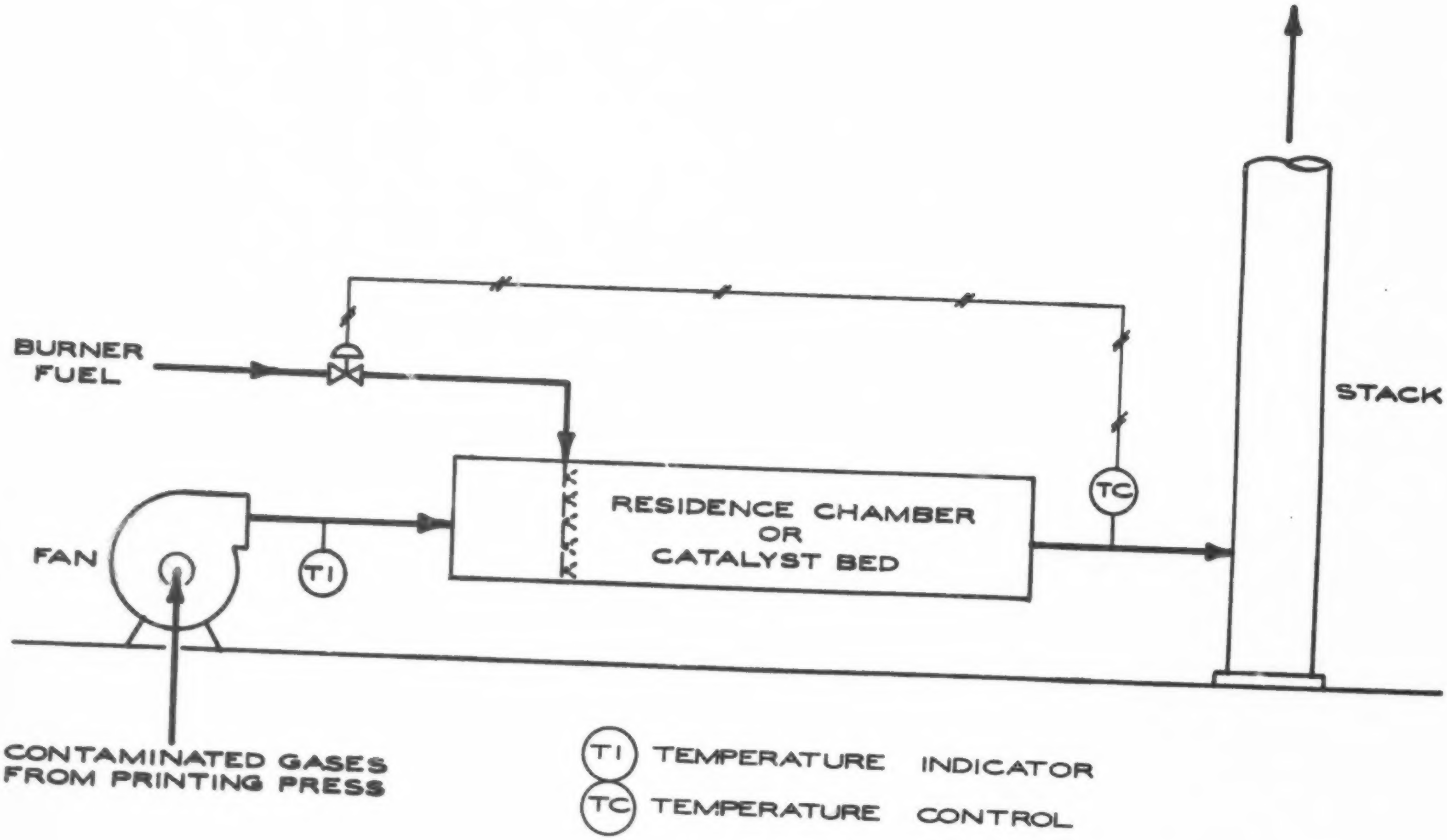

BURNER FUEL
FAN
RESIDENCE CHAMBER OR CATALYST BED
STACK
TC
TI
CONTAMINATED GASES FROM PRINTING PRESS
TI TEMPERATURE INDICATOR
TC TEMPERATURE CONTROL

TABLE 130

CATALYTIC INCINERATOR OPERATING CONDITIONS FOR WEB-OFFSET PRINTING SPECIFICATION (WITHOUT HEAT EXCHANGE)

Plant Size	Small	Large
Process Conditions		
Effluent Gas Temperature, °F	350	350
Effluent Gas Rate, SCFM	2,000	7,000
Effluent Gas Rate, ACFM	3,060	10,700
Hcbn Concentration, ppm(1)	500-2,000	500-2,000
Hcbn Emission Rate, lb/hr		
Maximum	10.1	35.4
Average	6.3	22.1
Heat of Combustion of Hcbn, Btu/lb(2)	19,500	19,500
Heating Value of Gas, Btu/SCF		
Maximum	1.5	1.5
Average	0.9	0.9
Incinerator Specifications		
Inlet Temperature, °F	350	350
Outlet Temperature, °F	1,100	1,100
Burner Duty, MMBtu/hr*	1.89	6.60

(1) Measured as methane equivalents
(2) Assumed as n-hexane

*To be filled in by supplier

Preceding page blank

TABLE 131

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CATALYTIC INCINERATORS FOR WEB-OFFSET PRINTING (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			3,060	10,700
°F			350	350
SCFM			2,000	7,000
Moisture Content, Vol. %			-	-
Effluent Contaminant Loading				
ppm			500-2,000	500-2,000
lb/hr avg			6.3	22.1
Btu/scf			0.9	0.9
Cleaned Gas Flow				
ACFM			5,970	20,900
°F			1,100	1,100
SCFM			2,031	7,105
Moisture Content, Vol. %			-	-
Cleaned Gas Contaminant Loading				
lb/hr			0.3	1.1
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			17,960	34,300
(2) Auxiliaries Cost			1,790	3,690
(a) Fan(s)				
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering				
(b) Foundations & Support			9,485	11,938
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost			29,235	49,928

FIGURE 101

CAPITAL COSTS FOR CATALYTIC INCINERATORS FOR WEB-OFFSET PRINTING (WITHOUT HEAT EXCHANGE)

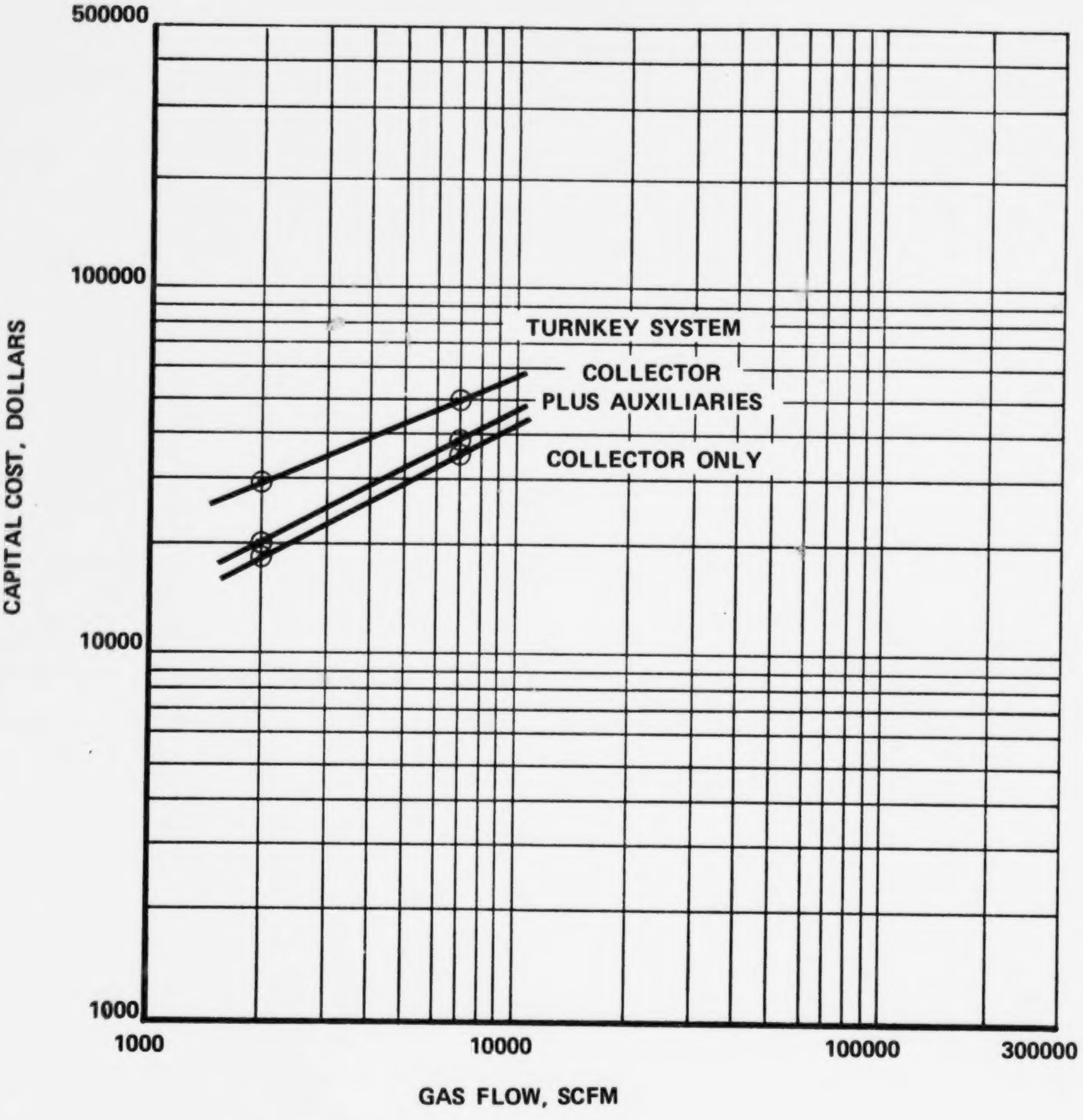


TABLE 132

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR CATALYTIC INCINERATORS FOR WEB-OFFSET PRINTING (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				4,000	4,000
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor				875	875
Maintenance					
Labor	$6/hr			510	510
Materials				110	135
Total Maintenance				620	645
Replacement Parts				943	2,950
Total Replacement Parts				943	2,950
Utilities					
Electric Power	$.011/kw-hr			423	1,369
Fuel	$.80/MMBtu			5,842	20,466
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				6,265	21,835
Total Direct Cost				8,703	26,305
Annualized Capital Charges				2,924	4,993
Total Annual Cost				11,627	31,298

FIGURE 102

ANNUAL COSTS FOR CATALYTIC INCINERATORS FOR WEB-OFFSET PRINTING (WITHOUT HEAT EXCHANGE)

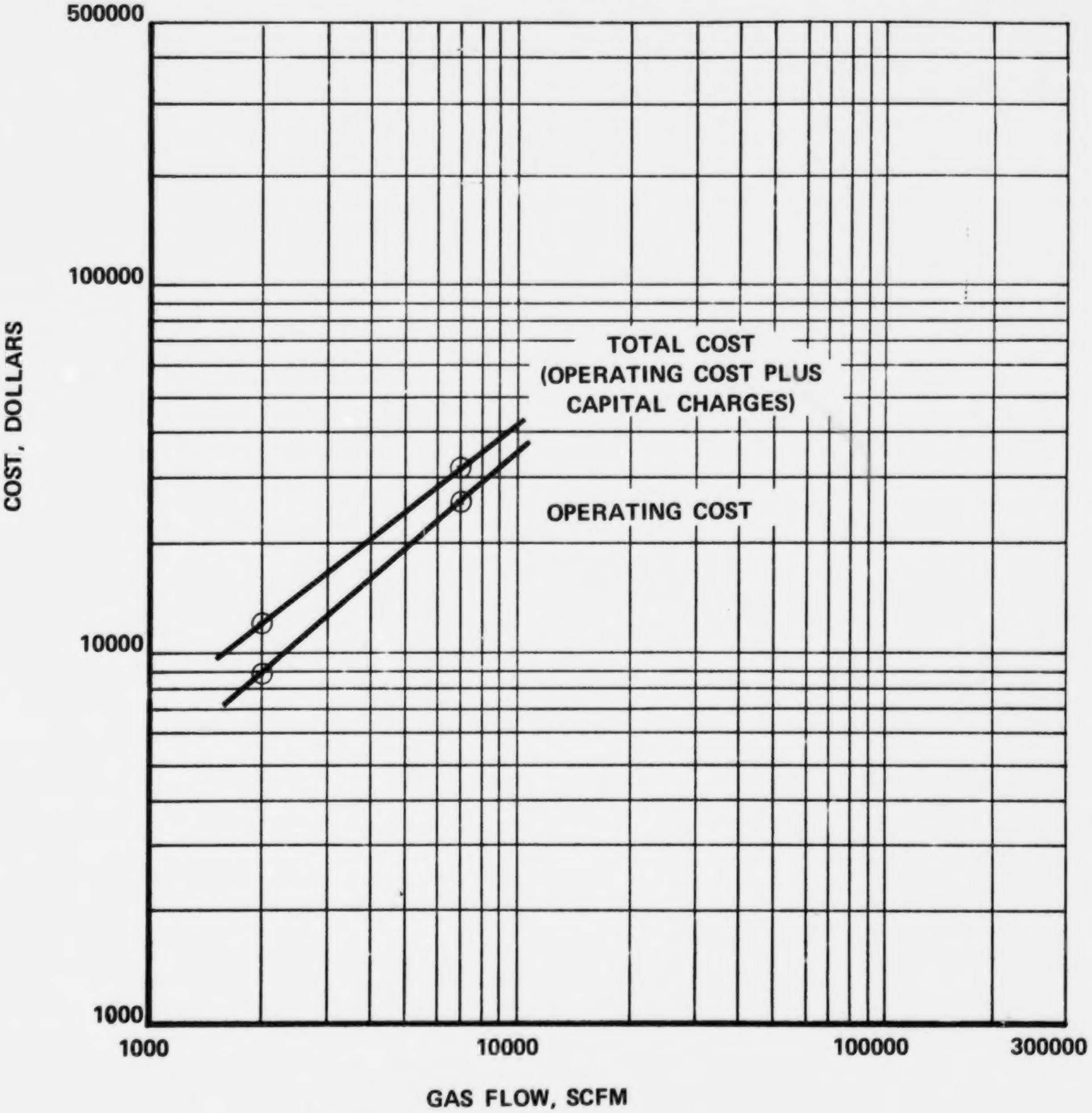


TABLE 133

THERMAL INCINERATOR PROCESS DESCRIPTION FOR METAL DECORATING SPECIFICATION

This specification describes the requirements of a thermal combustion system for the abatement of hydrocarbon emissions from metal decorating facilities. Schematic flow diagrams of the desired combustion systems are included in Figure 103 while processing conditions and specifications for small and large facilities are tabulated in Table 134.

The incinerators will be gas fired using natural gas available at a pressure of 1.0 psig and having a specific gravity of 0.6 and an upper heating value of 1,040 Btu/SCF.

The exhaust gas from the printing press will contain sufficient oxygen (greater than 16%) to allow firing of the burner without the addition of a combustion air system.

A fan equipped with a V-belt drive will be required at the incinerator inlet. The fan will have the capacity to overcome the pressure drop of the ductwork, burner, and heat exchange that may be incorporated in the system. The system's ductwork will be sized for a maximum ΔP *of 2 in. w.c. (hot).*

The incinerators will be supplied with suitable control panels, and all equipment will be designed for outdoor operation and to meet Factory Insurance Association's standards.

The cost estimates for each combustion system will include the following items:

Incinerator
Burner
10 ft stack
Controls
Control panel
Structural steel
Fuel gas piping
Electrical
Ductwork
Insulation
Fan
Fan motor
Two day start-up service
Dampers (when included)

All items with exception of the incinerator and burner will be considered auxiliaries.

The incinerators shall be located on the roof of the printing facility, and no modification of the building structure is required. All utilities are available within 30 ft of the control cabinet, motor, and burner.

FIGURE 103

SCHEMATIC FLOW DIAGRAM OF THERMAL AND CATALYTIC COMBUSTION WITHOUT HEAT EXCHANGE FOR WEB-OFF SET AND METAL DECORATING

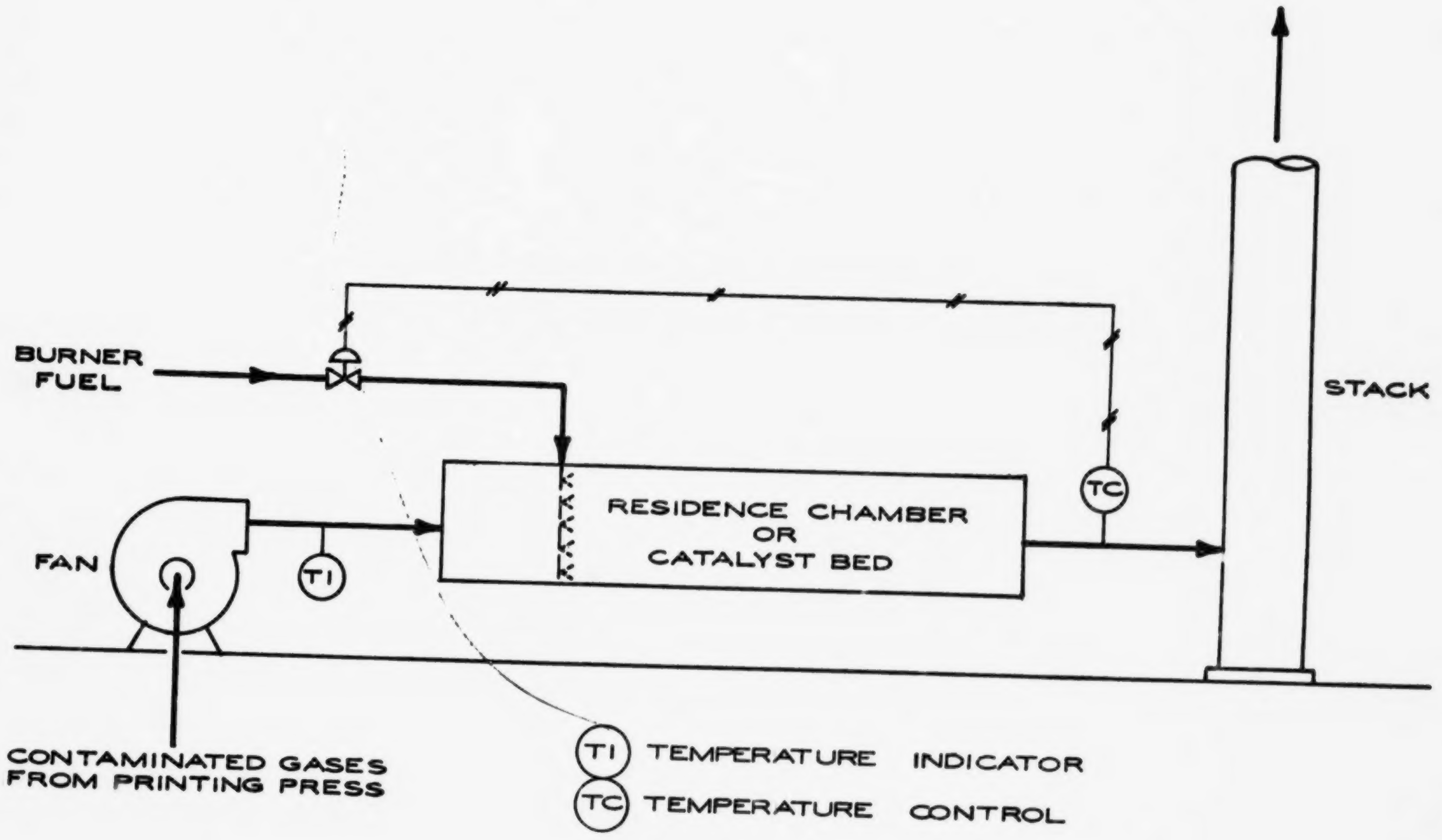

BURNER FUEL
FAN
TI
TC
RESIDENCE CHAMBER OR CATALYST BED
STACK
CONTAMINATED GASES FROM PRINTING PRESS
TI TEMPERATURE INDICATOR
TC TEMPERATURE CONTROL

TABLE 134

THERMAL INCINERATOR OPERATING CONDITIONS FOR METAL DECORATING SPECIFICATION (WITHOUT HEAT EXCHANGE)

Plant Size	*Small*	*Large*
Process Conditions		
Effluent Gas Temperature, °F	350	350
Effluent Gas Rate, SCFM	4,000	10,000
Effluent Gas Rate, ACFM	6,100	15,250
Hcbn Concentration, % LEL	15	15
Hcbn Emission Rate, lb/hr	97	242
Heating Value of Gas, Btu/SCF	7.4	7.4
Heat of Combustion of Hcbn, Btu/lb	18,400	18,400
Incinerator Specifications		
Inlet Gas Temperature, °F	350	350
Outlet Gas Temperature, °F	1,500	1,500
Residence Time @ Temperature, sec	0.6	0.6
*Burner Duty, MMBtu/hr**	4.21	11.07

**To be filled in by supplier*

Preceding page blank

TABLE 135

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR METAL DECORATING (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			6,100	15,250
°F			350	350
SCFM			4,000	10,000
Moisture Content, Vol. %			-	-
Effluent Hydrocarbon				
%LEL			15	15
lb/hr			97	242
Btu/scf			7.4	7.4
Cleaned Gas Flow				
ACFM			15,070	37,684
°F			1,500	1,500
SCFM			4,075	10,190
Moisture Content, Vol. %			-	-
Cleaned Gas Contaminant Loading				
lb/hr			<4.9	<12.1
Cleaning Efficiency, %			>95	>95
(1) Gas Cleaning Device Cost			16,775	27,800
(2) Auxiliaries Cost				
(a) Fan(s)				
(b) Pump(s)			2,338	3,995
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			2,850	3,500
(b) Foundations & Support			2,450	2,800
Ductwork			2,750	4,000
Stack			825	1,325
Electrical			750	1,250
Piping			250	425
Insulation				
Painting				
Supervision			950	950
Startup			450	450
Performance Test			1,750	1,750
Other			4,525	5,925
(4) Total Cost			36,663	54,170

FIGURE 104

CAPITAL COSTS FOR THERMAL INCINERATORS
FOR METAL DECORATING
(WITHOUT HEAT EXCHANGE)

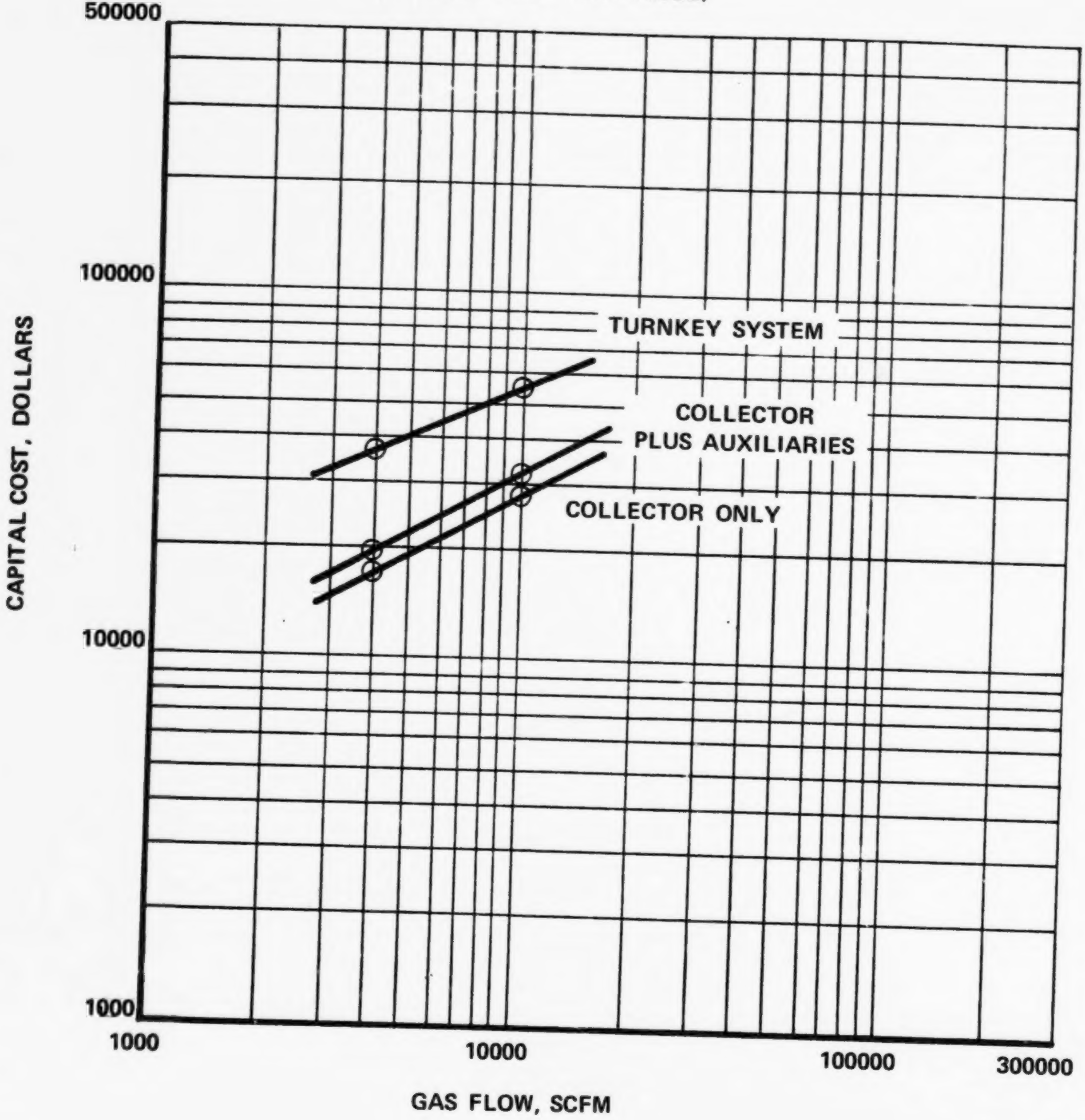


TABLE 136

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR THERMAL INCINERATORS FOR METAL DECORATING (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				4,000	4,000
Operating Labor (if any)					
Operator	$6/hr			1,500	1,500
Supervisor					
Total Operating Labor				1,500	1,500
Maintenance					
Labor	$6/hr			336	336
Materials				135	160
Total Maintenance				471	496
Replacement Parts				125	125
Total Replacement Parts				125	125
Utilities					
Electric Power	$.011/kw-hr			510	1,097
Fuel	$.80/MMBtu			14,488	37,044
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				14,998	38,141
Total Direct Cost				17,094	40,262
Annualized Capital Charges				3,666	5,417
Total Annual Cost				20,760	45,679

FIGURE 105

ANNUAL COSTS FOR THERMAL INCINERATORS FOR METAL DECORATING (WITHOUT HEAT EXCHANGE)

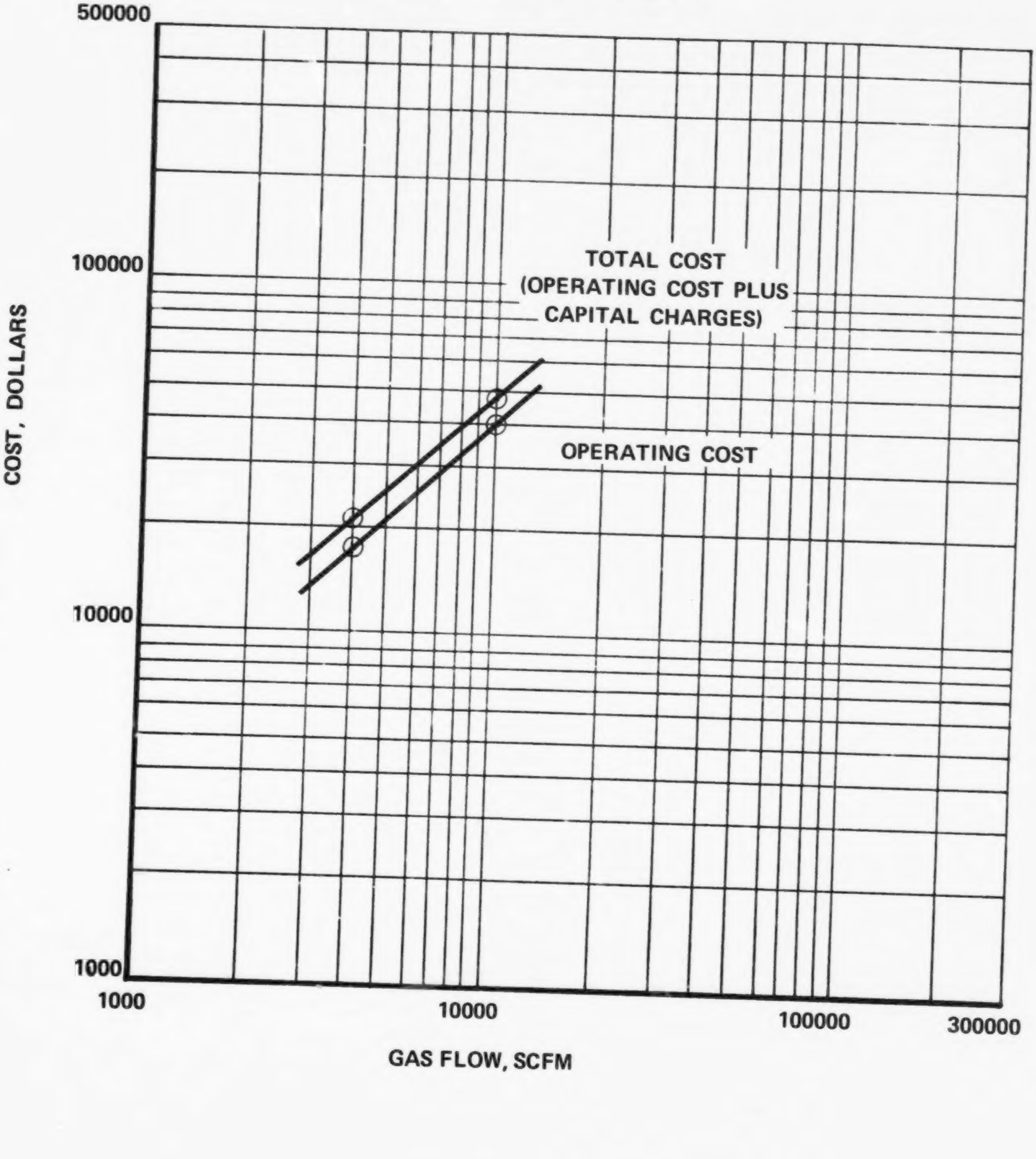


TABLE 137

THERMAL INCINERATOR PROCESS DESCRIPTION FOR METAL DECORATING SPECIFICATION

This specification describes the requirements of a thermal combustion system for the abatement of hydrocarbon emissions from metal decorating facilities. Schematic flow diagrams of the desired combustion systems are included in Figure 106 while processing conditions and specifications for small and large facilities are tabulated in Table 138.

The incinerators will be gas fired using natural gas available at a pressure of 1.0 psig and having a specific gravity of 0.6 and an upper heating value of 1,040 Btu/SCF.

The exhaust gas from the printing press will contain sufficient oxygen (greater than 16%) to allow firing of the burner without the addition of a combustion air system.

A fan equipped with a V-belt drive will be required at the incinerator inlet. The fan will have the capacity to overcome the pressure drop of the ductwork, burner, and heat exchange that may be incorporated in the system. The system's ductwork will be sized for a maximum ΔP of 2 in. w.c. (hot).

The heat exchanger shall be a parallel flow shell and tube exchanger and designed to operate at an incinerator outlet temperature of 1,500°F. Maximum exchanger pressure drop, both sides, shall not exceed 6.0 in. w.c. (hot), and dirty gas shall flow through the tube side. Approximate thermal efficiencies of the exchangers are tabulated in the specification sheets.

The incinerators will be supplied with suitable control panels, and all equipment will be designed for outdoor operation and to meet Factory Insurance Association's standards.

The cost estimates for each combustion system will include the following items:

Incinerator
Burner
10 ft stack
Controls
Control panel
Structural steel
Fuel gas piping
Electrical
Ductwork
Insulation
Fan
Fan motor
Two day start-up service
Heat exchanger
Dampers (when included)

All items with exception of the incinerator, burner, and heat exchanger will be considered auxiliaries.

The incinerators shall be located on the roof of the printing facility, and no modification of the building structure is required. All utilities are available within 30 ft of the control cabinet, motor, and burner.

FIGURE 106

SCHEMATIC FLOW DIAGRAM OF THERMAL AND CATALYTIC COMBUSTION WITH HEAT EXCHANGE FOR WEB-OFFSET AND METAL DECORATING

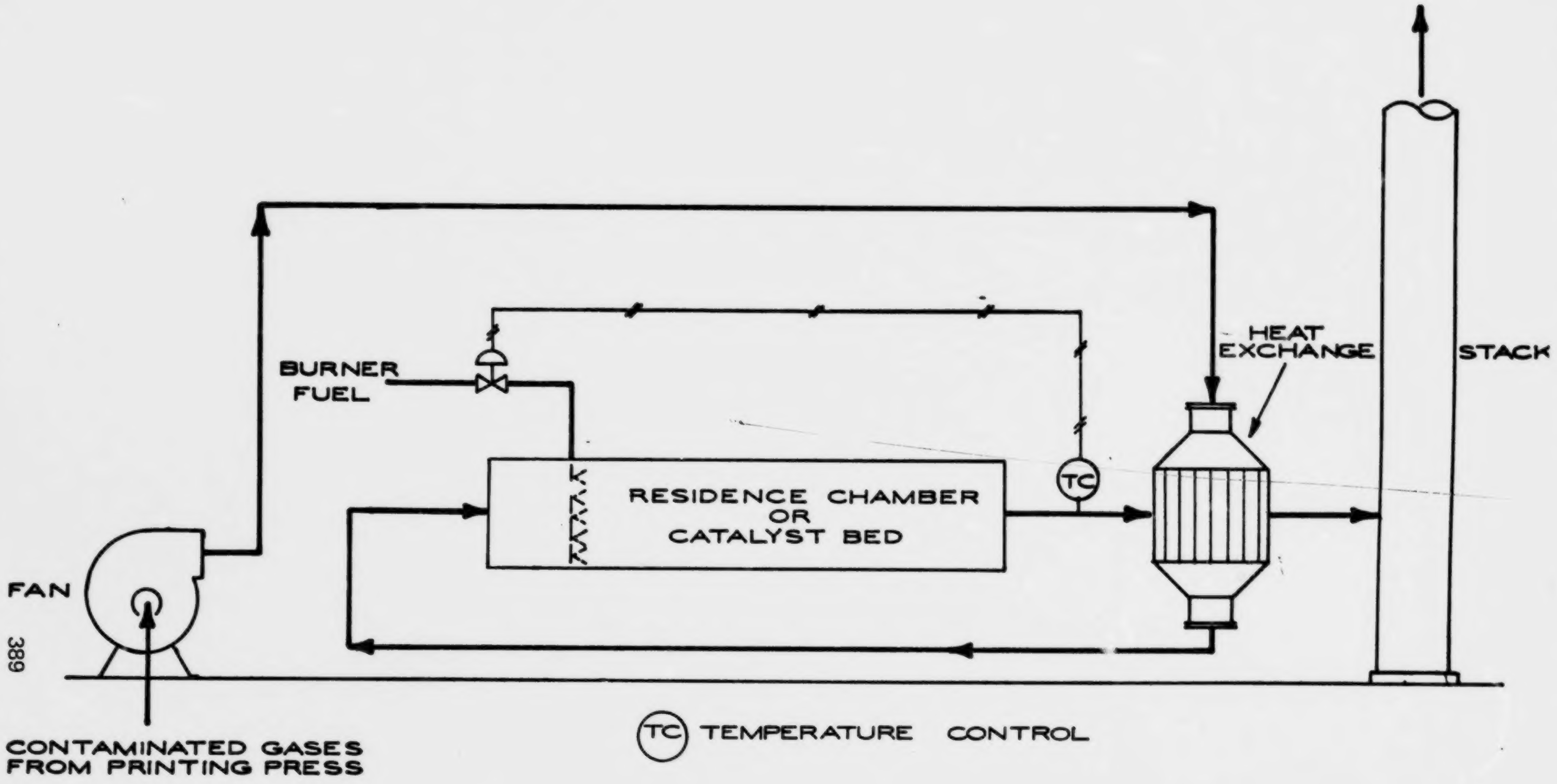

HEAT EXCHANGE
STACK
BURNER FUEL
RESIDENCE CHAMBER OR CATALYST BED
TC
FAN
CONTAMINATED GASES FROM PRINTING PRESS
TC TEMPERATURE CONTROL

TABLE 138

THERMAL INCINERATOR OPERATING CONDITIONS FOR METAL DECORATING SPECIFICATION (WITH HEAT EXCHANGE)

Plant Size	*Small*	*Large*
Process Conditions		
Effluent Gas Temperature, °F	*350*	*350*
Effluent Gas Rate, SCFM	*4,000*	*10,000*
Effluent Gas Rate, ACFM	*6,100*	*15,250*
Hcbn Concentration, % LEL	*15*	*15*
Hcbn Emission Rate, lb/hr	*97*	*242*
Heating Value of Gas, Btu/SCF	*7.4*	*7.4*
Heat of Combustion of Hcbn, Btu/lb	*18,400*	*18,400*
Incinerator Specifications		
Residence Time, sec @ Temperature	*0.6*	*0.6*
Inlet Tubeside Temperature, °F	*350*	*350*
Unit Outlet Temperature, °F	*1,500*	*1,500*
Thermal Efficiency	*≃42*	*≃42*
*Burner Duty, MMBtu/hr**	*1.83*	*3.49*

**To be filled in by supplier*

Preceding page blank

TABLE 139

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			6,100	15,250
°F			350	350
SCFM			4,000	10,000
Moisture Content, Vol. %			-	-
Effluent Hydrocarbon				
%LEL			15	15
lb/hr			97	242
Btu/scf			7.4	7.4
Cleaned Gas Flow				
ACFM			15,070	37,684
°F			1,500	1,500
SCFM			4,075	10,190
Moisture Content, Vol. %			-	-
Cleaned Gas Contaminant Loading				
ppm				
lb/hr			<4.9	<12.1
Cleaning Efficiency, %			>95	>95
(1) Gas Cleaning Device Cost			32,500	39,575
(2) Auxiliaries Cost				
(a) Fan(s)				
(b) Pump(s)				
(c) Damper(s)			2,788	5,680
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			3,100	3,750
(b) Foundations & Support			2,750	4,000
Ductwork			2,750	4,000
Stack			800	1,300
Electrical			875	1,625
Piping			250	400
Insulation				
Painting				
Supervision			1,050	1,050
Startup			450	450
Performance Test			1,750	1,750
Other			5,860	5,960
(4) Total Cost			54,923	69,540

FIGURE 107

CAPITAL COSTS FOR THERMAL INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

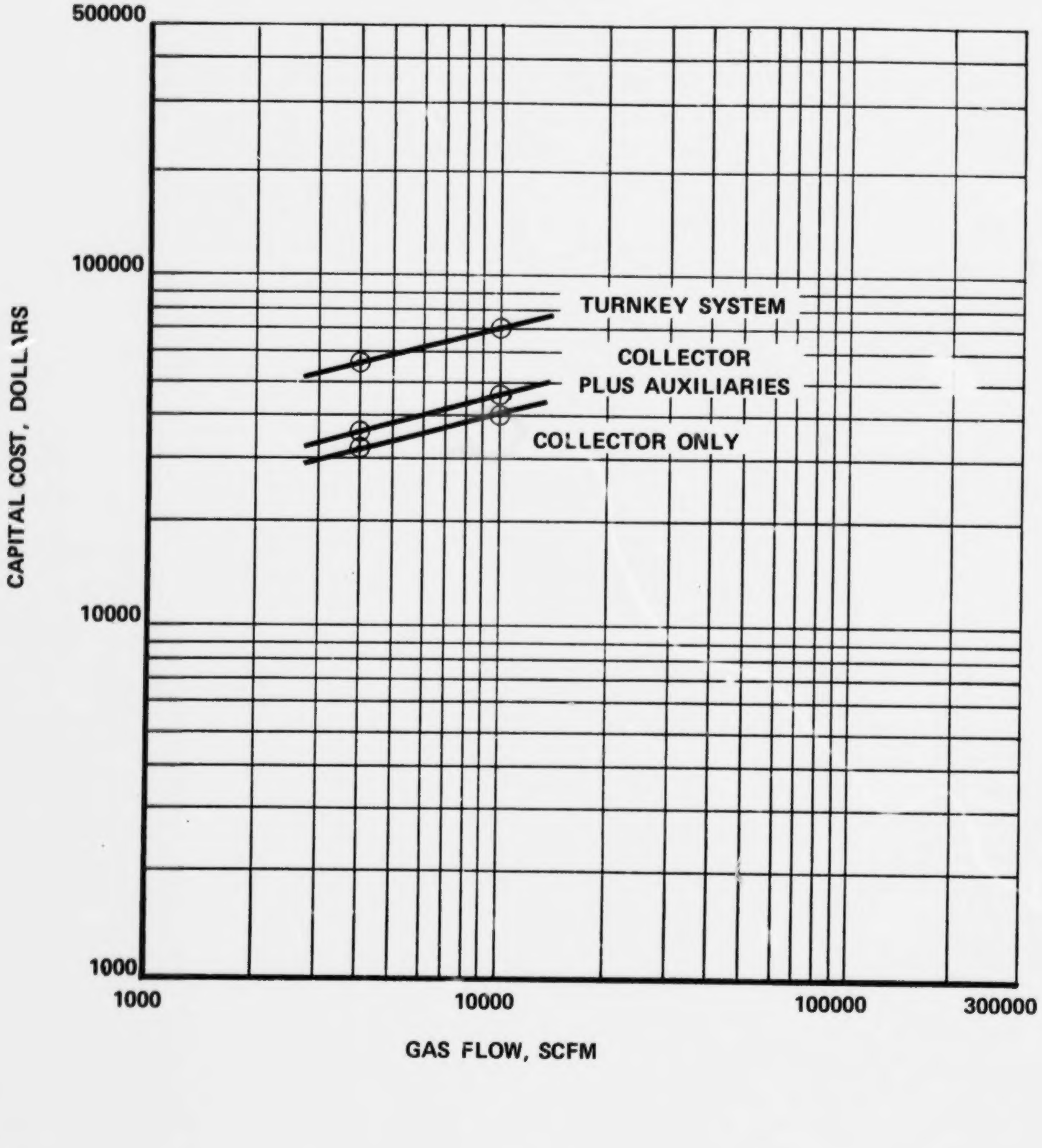


TABLE 140

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR THERMAL INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				4,000	4,000
Operating Labor (if any)					
Operator	$6/hr			1,575	1,575
Supervisor					
Total Operating Labor				1,575	1,575
Maintenance					
Labor	$6/hr			360	360
Materials				135	160
Total Maintenance				495	520
Replacement Parts				125	125
Total Replacement Parts				125	125
Utilities					
Electric Power	$.011/kw-hr			1,043	2,149
Fuel	$.80/MMBtu			5,479	12,036
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				6,522	14,185
Total Direct Cost				8,717	16,405
Annualized Capital Charges				5,492	6,954
Total Annual Cost				14,209	23,359

FIGURE 108

ANNUAL COSTS FOR THERMAL INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

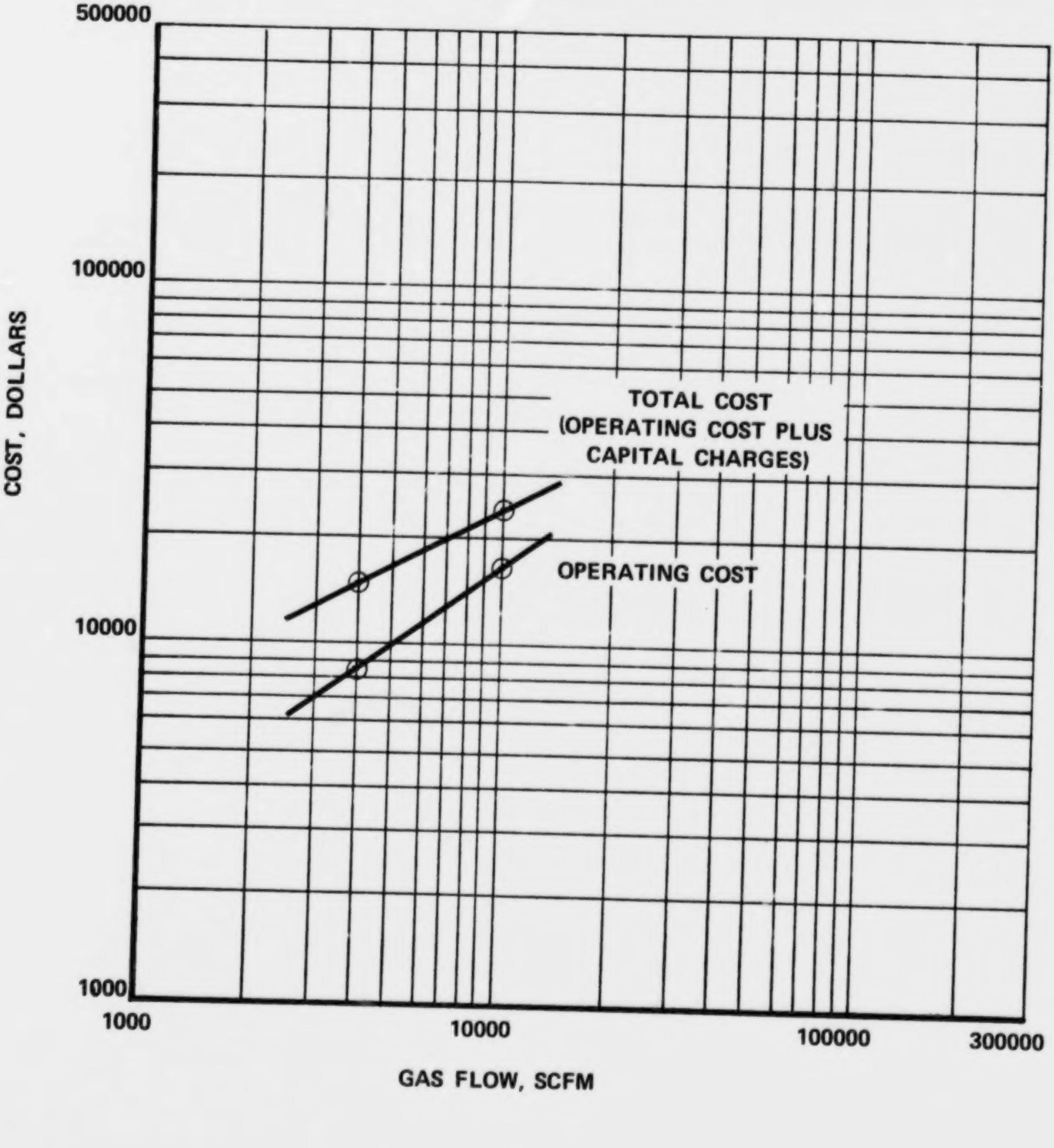


TABLE 141

CATALYTIC INCINERATOR PROCESS DESCRIPTION FOR METAL DECORATING SPECIFICATION

This specification describes the requirements of catalytic combustion systems for the abatement of hydrocarbon emissions from metal decorating facilities. Schematic flow diagrams of the desired combustion systems are included in Figure 109 while processing conditions and specifications for small and large facilities are tabulated in Table 142.

The incinerators will be gas fired using natural gas available at a pressure of 1.0 psig and having a specific gravity of 0.6 and an upper heating value of 1,040 Btu/SCF.

The exhaust gas from the printing press will contain sufficient oxygen (greater than 16%) to allow firing of the burner without the addition of a combustion air system.

A fan equipped with a V-belt drive will be required at the incinerator inlet. The fan will have the capacity to overcome the pressure drop of the ductwork, burner, and any heat exchange that may be incorporated in the system. The system's ductwork will be sized for a maximum ΔP of 2 in. w.c. (hot).

The incinerators will be supplied with suitable control panels, and all equipment will be designed for outdoor operation and to meet Factory Insurance Association's standards.

The cost estimates for each combustion system shall include the following items:

Incinerator
Burner
10 ft stack
Controls
Control panel
Structural steel
Fuel gas piping
Electrical
Ductwork
Insulation
Fan
Fan motor
Two day start-up service
Dampers (when included)

All items with exception of the incinerator and burner will be considered auxiliaries.

The incinerators will be located on the roof of the printing facility, and no modification of the building structure is required. All utilities are available within 30 ft of the control cabinet, motor, and burner.

FIGURE 109

SCHEMATIC FLOW DIAGRAM OF THERMAL AND CATALYTIC COMBUSTION WITHOUT HEAT EXCHANGE FOR WEB OFFSET AND METAL DECORATING

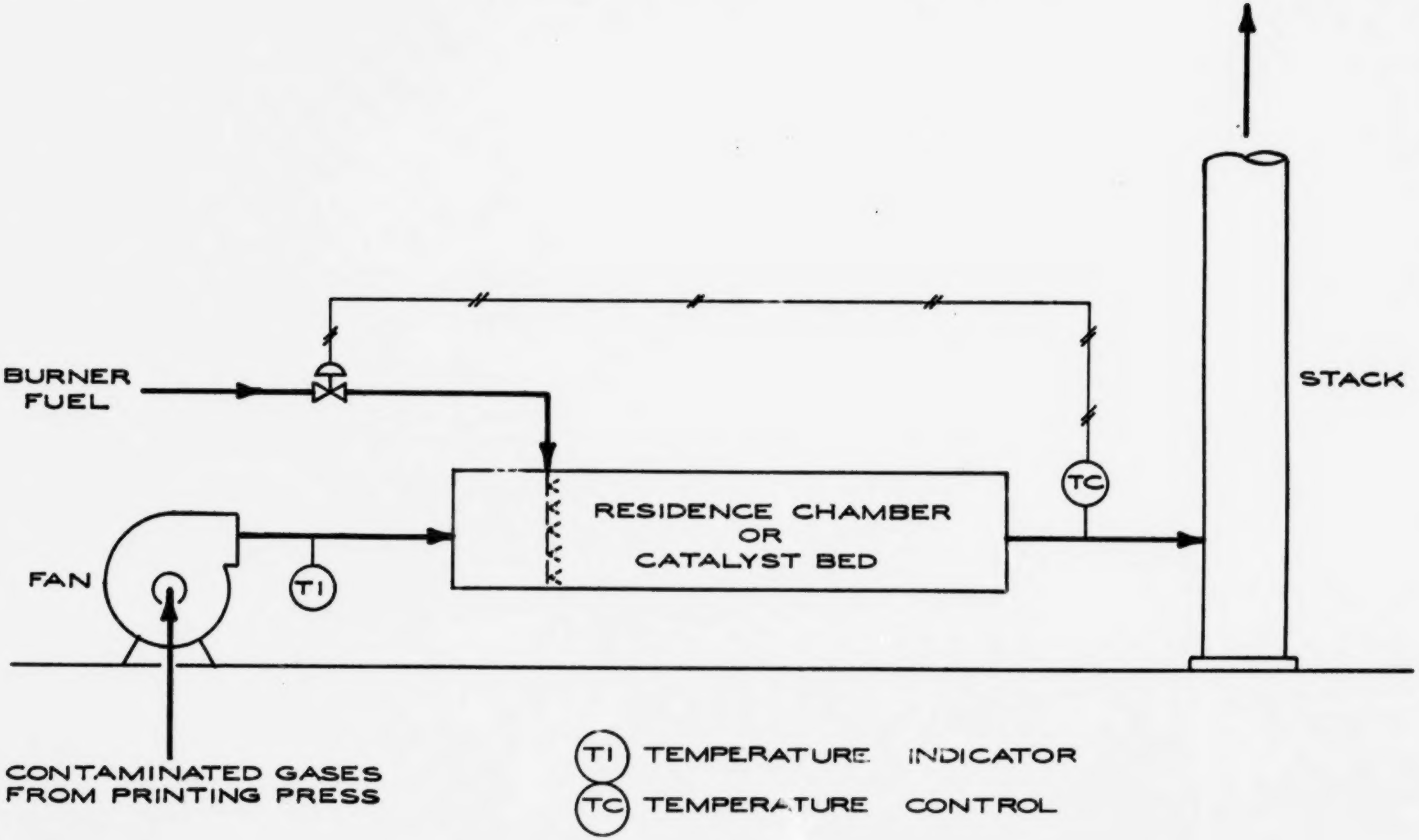

BURNER FUEL
FAN
RESIDENCE CHAMBER OR CATALYST BED
TC
TI
STACK
CONTAMINATED GASES FROM PRINTING PRESS
TI TEMPERATURE INDICATOR
TC TEMPERATURE CONTROL

TABLE 142

CATALYTIC INCINERATOR OPERATING CONDITIONS FOR METAL DECORATING SPECIFICATION (WITHOUT HEAT EXCHANGE)

Plant Size	*Small*	*Large*
Process Conditions		
Effluent Gas Temperature, °F	*350*	*350*
Effluent Gas Rate, SCFM	*4,000*	*10,000*
Effluent Gas Rate, ACFM	*6,100*	*15,250*
Hcbn Concentration, % LEL	*15*	*15*
Hcbn Emission Rate, lbs/hr	*97*	*242*
Heating Value of Gas, Btu/SCF	*7.4*	*7.4*
Heat of Combustion of Hcbn, Btu/lb	*18,400*	*18,400*
Incinerator Specifications		
Inlet Gas Temperature, °F	*350*	*350*
Outlet Gas Temperature, °F	*1,100*	*1,100*
**Burner Duty, MMBtu/hr*	*1.97*	*4.92*

**To be filled in by supplier*

Preceding page blank

TABLE 143

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITHOUT HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			6,100	15,250
°F			350	350
SCFM				
Moisture Content, Vol. %			4,000	10,000
Effluent Hydrocarbon				
% LEL			15	15
lb/hr			97	242
Btu/scf			7.4	7.4
Cleaned Gas Flow				
ACFM			11,860	29,650
°F			1,100	1,100
SCFM			4,033	10,083
Moisture Content, Vol. %				
Cleaned Gas Contaminant Loading				
lb/hr			4.9	12.1
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			26,010	42,560
(2) Auxiliaries Cost				
(a) Fan(s)			2,270	4,940
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			10,565	14,775
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost			38,845	62,275

FIGURE 110

CAPITAL COSTS FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITHOUT HEAT EXCHANGE)

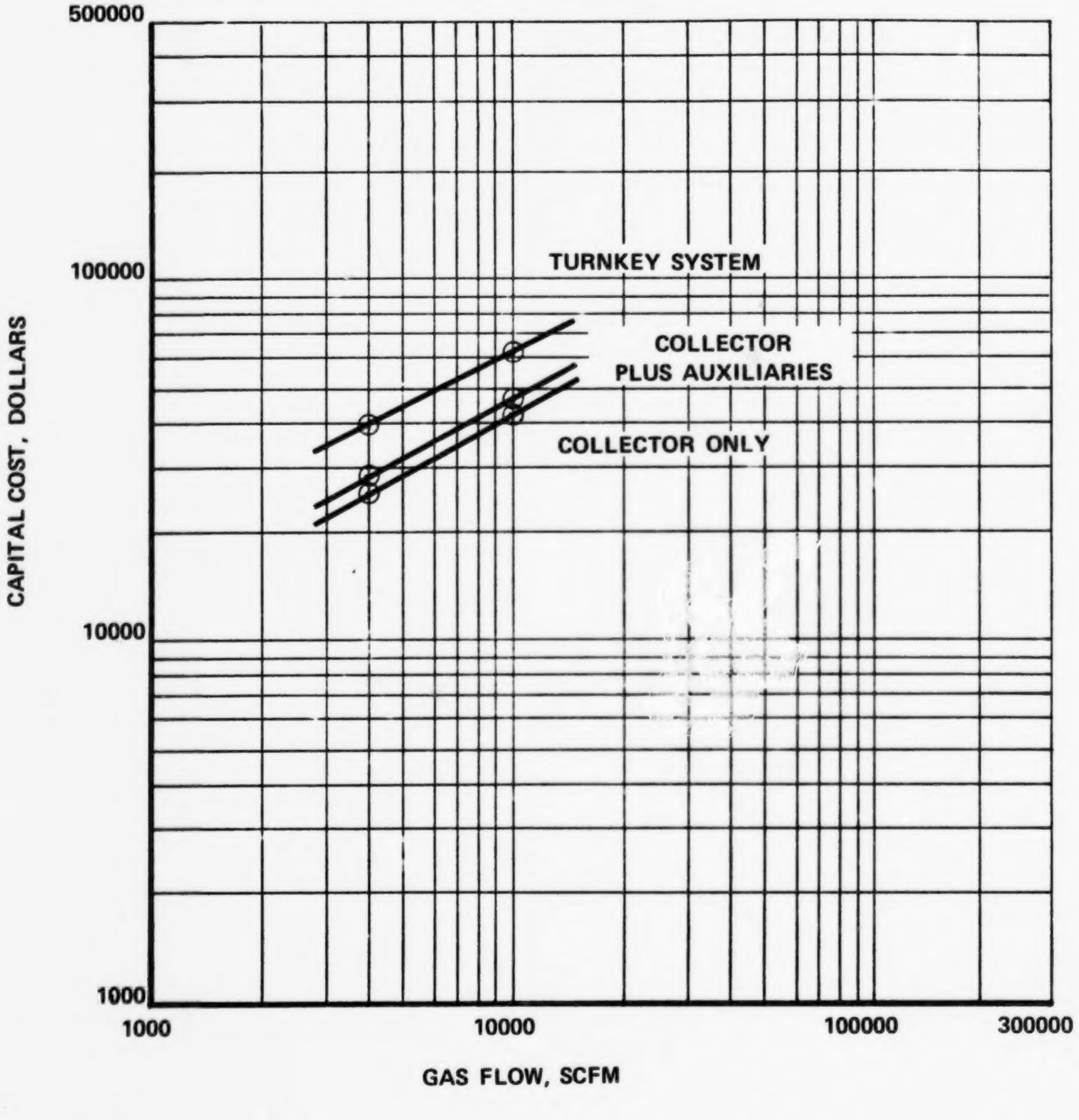


TABLE 144

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITHOUT HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				4,000	4,000
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor				875	875
Maintenance					
Labor	$6/hr			510	510
Materials				110	135
Total Maintenance				620	645
Replacement Parts					
Total Replacement Parts				1,708	4,175
Utilities					
Electric Power	$.011/kw-hr			824	2,058
Fuel	$.80/MMBtu			6,490	16,225
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				7,314	18,283
Total Direct Cost				10,517	23,978
Annualized Capital Charges				3,885	6,228
Total Annual Cost				14,402	30,206

FIGURE 111

ANNUAL COSTS FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITHOUT HEAT EXCHANGE)

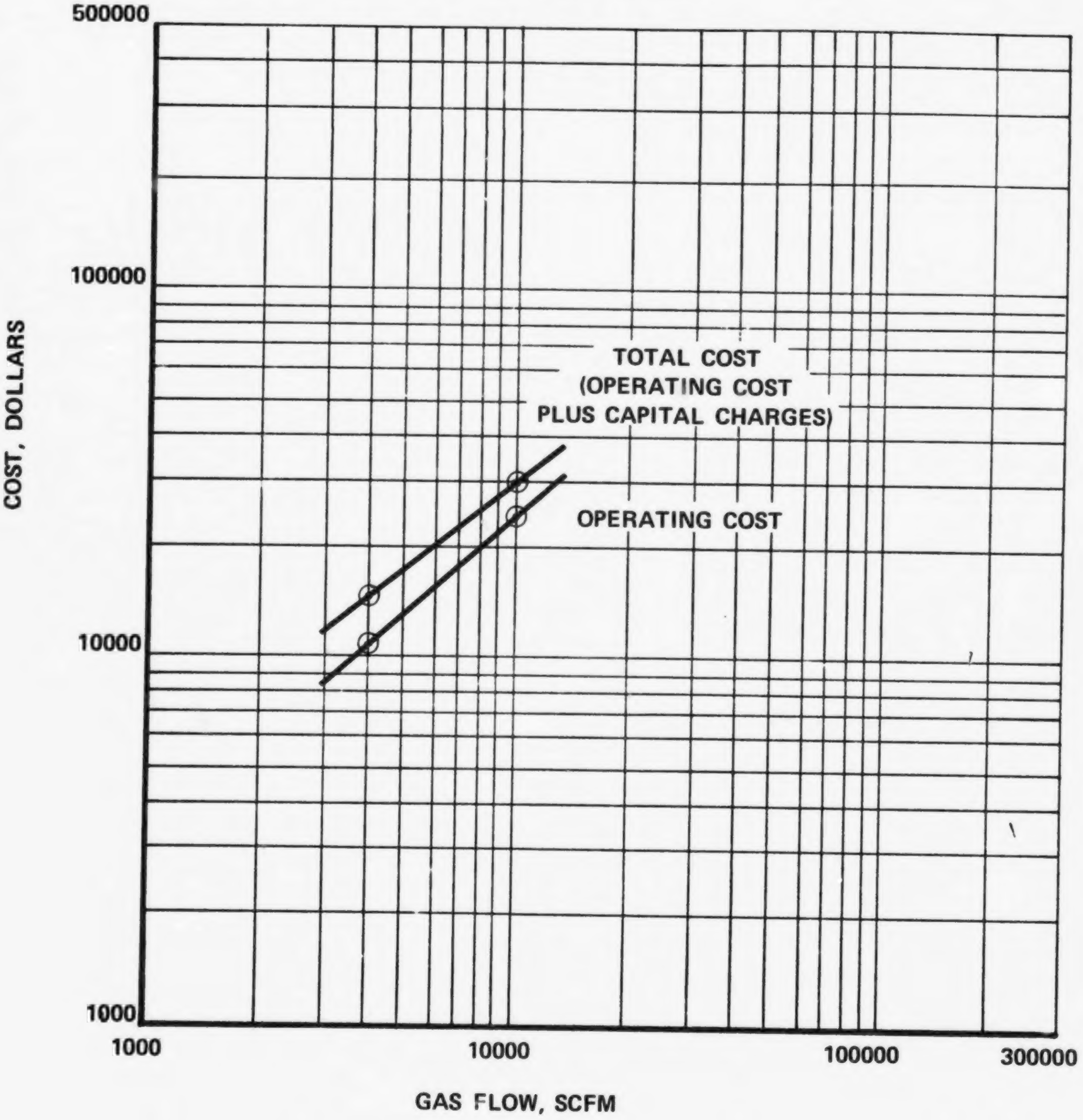


TABLE 145

CATALYTIC INCINERATOR PROCESS DESCRIPTION FOR METAL DECORATING SPECIFICATION

This specification describes the requirements of catalytic combustion systems for the abatement of hydrocarbon emissions from metal decorating facilities. Schematic flow diagrams of the desired combustion systems ary included in Figure 112 while processing conditions and specifications for small and large facilities are tabulated in Table 146.

The incinerators will be gas fired using natural gas available at a pressure of 1.0 psig and having a specific gravity of 0.6 and an upper heating value of 1,040 Btu/SCF.

The exhaust gas from the printing press will contain sufficient oxygen (greater than 16%) to allow firing of the burner without the addition of a combustion air system.

A fan equipped with a V-belt drive will be required at the incinerator inlet. The fan will have the capacity to overcome the pressure drop of the ductwork, burner, and any heat exchange that may be incorporated in the system. The system's ductwork will be sized for a maximum ΔP of 2 in. w.c. (hot).

The heat exchanger will be a parallel flow shell and tube exchanger and designed to operate at an outlet temperature of 1,100°F. Maximum exchanger pressure drop, both sides, will not exceed 6.0 in. w.c. (hot), and dirty gas will flow through the tube side. Approximate thermal efficiencies of the exchangers are tabulated in the specification sheets.

The incinerators will be supplied with suitable control panels, and all equipment will be designed for outdoor operation and to meet Factory Insurance Association's standards.

The cost estimates for each combustion system shall include the following items:

Incinerator
Burner
10 ft stack
Controls
Control panel
Structural steel
Fuel gas piping
Electrical
Ductwork
Insulation
Fan
Fan motor
Two day start-up service
Heat exchanger
Dampers (when included)

All items with exception of the incinerator, burner, and heat exchanger will be considered auxiliaries.

The incinerators will be located on the roof of the printing facility, and no modification of the building structure is required. All utilities are available within 30 ft of the control cabinet, motor, and burner.

FIGURE 112

SCHEMATIC FLOW DIAGRAM OF
THERMAL AND CATALYTIC
COMBUSTION WITH
HEAT EXCHANGE FOR
WEB-OFF SET AND METAL DECORATING

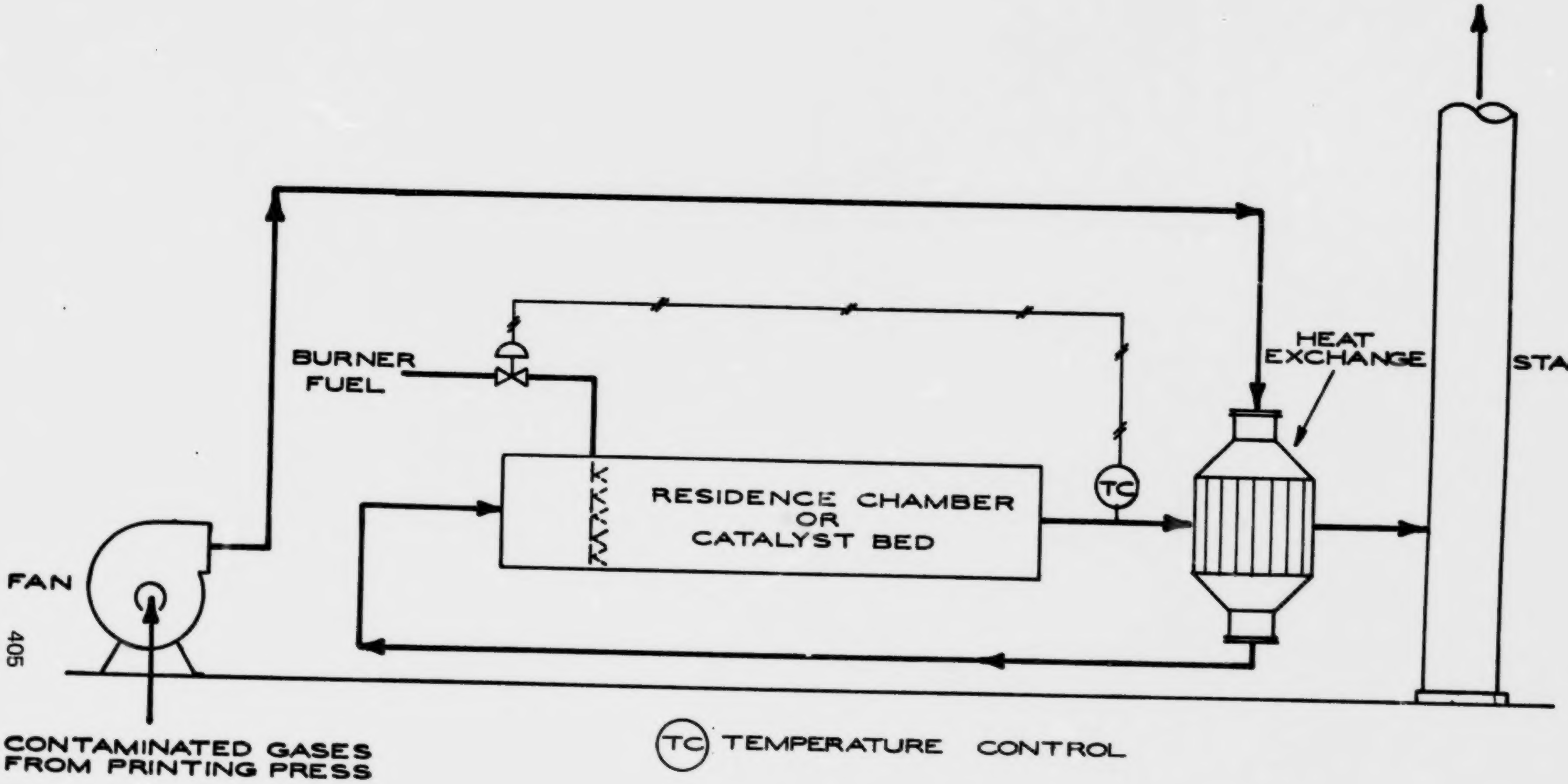

BURNER
FUEL
HEAT
EXCHANGE
STACK
TC
RESIDENCE CHAMBER
OR
CATALYST BED
FAN
CONTAMINATED GASES
FROM PRINTING PRESS
TC TEMPERATURE CONTROL

TABLE 146

CATALYTIC INCINERATOR OPERATING CONDITIONS FOR METAL DECORATING SPECIFICATION (WITH HEAT EXCHANGE)

Plant Size	*Small*	*Large*
Process Conditions		
Effluent Gas Temperature, °F	350	350
Effluent Gas Rate, SCFM	4,000	10,000
Effluent Gas Rate, ACFM	6,100	15,250
Hcbn Concentration, % LEL	15	15
Hcbn Emission Rate, lb/hr	97	242
Heating Value of Gas, Btu/SCF	7.4	7.4
Heat of Combustion of Hcbn, Btu/lb	18,400	18,400
Incinerator Specifications		
Inlet Tubeside Temperature, °F	350	350
Unit Outlet Temperature, °F	1,100	1,100
Burner Δ T from Fuel Gas, °F	25°F (min)	25°F (min)
Thermal Efficiency	*To obtain min burner Δ T*	
**Duty, MMBtu/hr*	1.71	4.28
**Burner Duty, MMBtu/hr*	1.97	4.92
**Overall Heat Transfer Coef., U*	4.5	4.5
**Tube Surface Area, ft^2*	1,100	2,740

**To be filled in by supplier*

Preceding page blank

TABLE 147

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			6,100	15,250
°F			350	350
SCFM			4,000	10,000
Moisture Content, Vol. %			-	-
Effluent Hydrocarbon				
%LEL			15	15
lb/hr			97	242
Btu/scf			7.4	7.4
Cleaned Gas Flow				
ACFM			11,800	29,400
°F			1,100	1,100
SCFM			4,000	10,000
Moisture Content, Vol. %			-	-
Cleaned Gas Contaminant Loading				
			-	-
lb/hr			4.85	12.1
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			36,160	55,950
(2) Auxiliaries Cost				
(a) Fan(s)				
(b) Pump(s)				
(c) Damper(s)			3,053	6,635
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering				
(b) Foundations & Support				
Ductwork			11,915	16,275
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost			51,128	78,860

FIGURE 113

CAPITAL COSTS FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

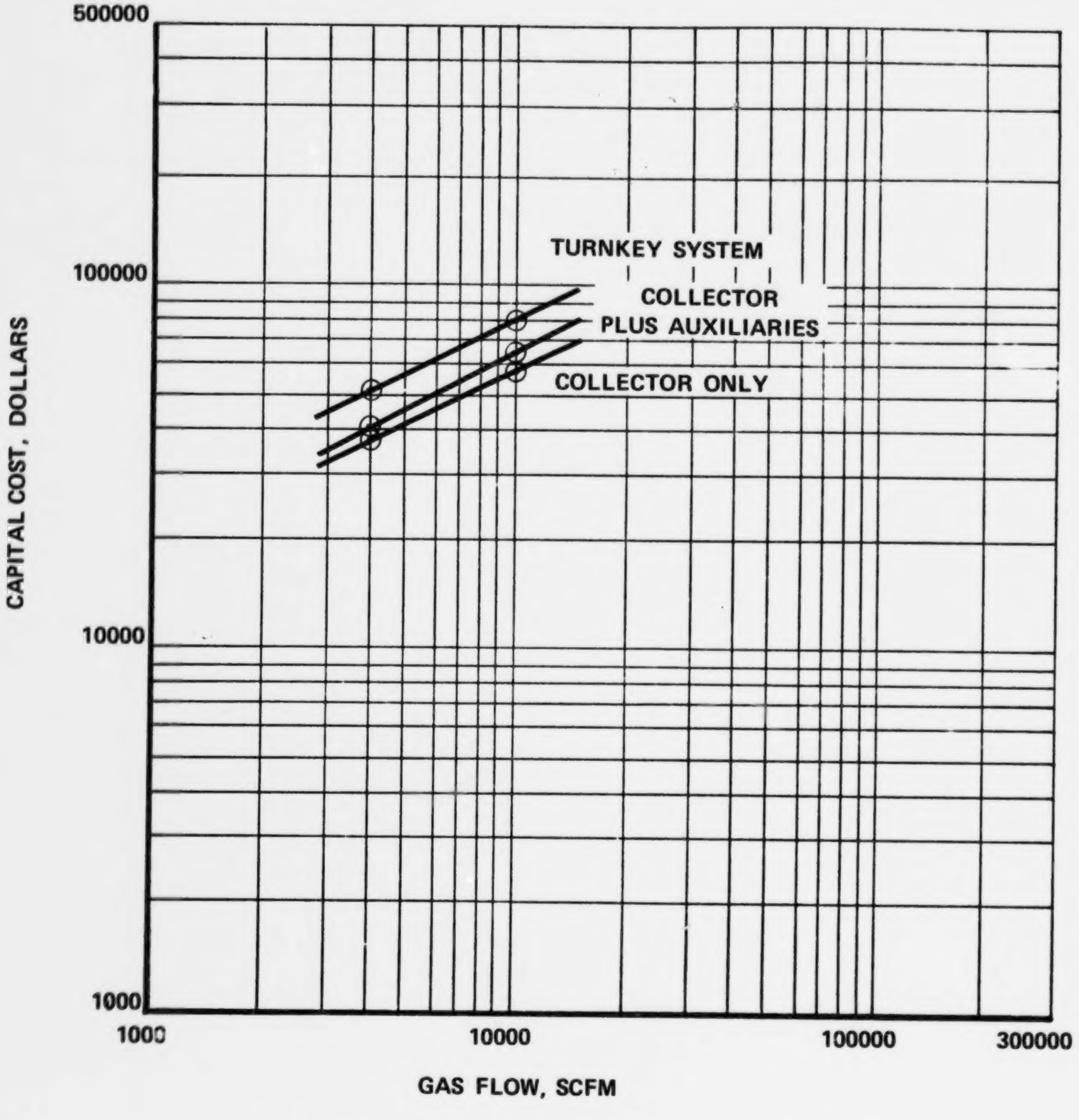


TABLE 148

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				4,000	4,000
Operating Labor (if any)					
Operator	$6/hr				
Supervisor	$8/hr				
Total Operating Labor				875	875
Maintenance					
Labor				510	510
Materials				110	135
Total Maintenance				620	645
Replacement Parts					
Total Replacement Parts				1,708	4,175
Utilities					
Electric Power	$.011/kw-hr			1,314	2,891
Fuel	$.80/MMBtu			266	657
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				1,580	3,548
Total Direct Cost				4,783	9,243
Annualized Capital Charges				5,113	7,886
Total Annual Cost				9,896	17,129

FIGURE 114

ANNUAL COSTS FOR CATALYTIC INCINERATORS FOR METAL DECORATING (WITH HEAT EXCHANGE)

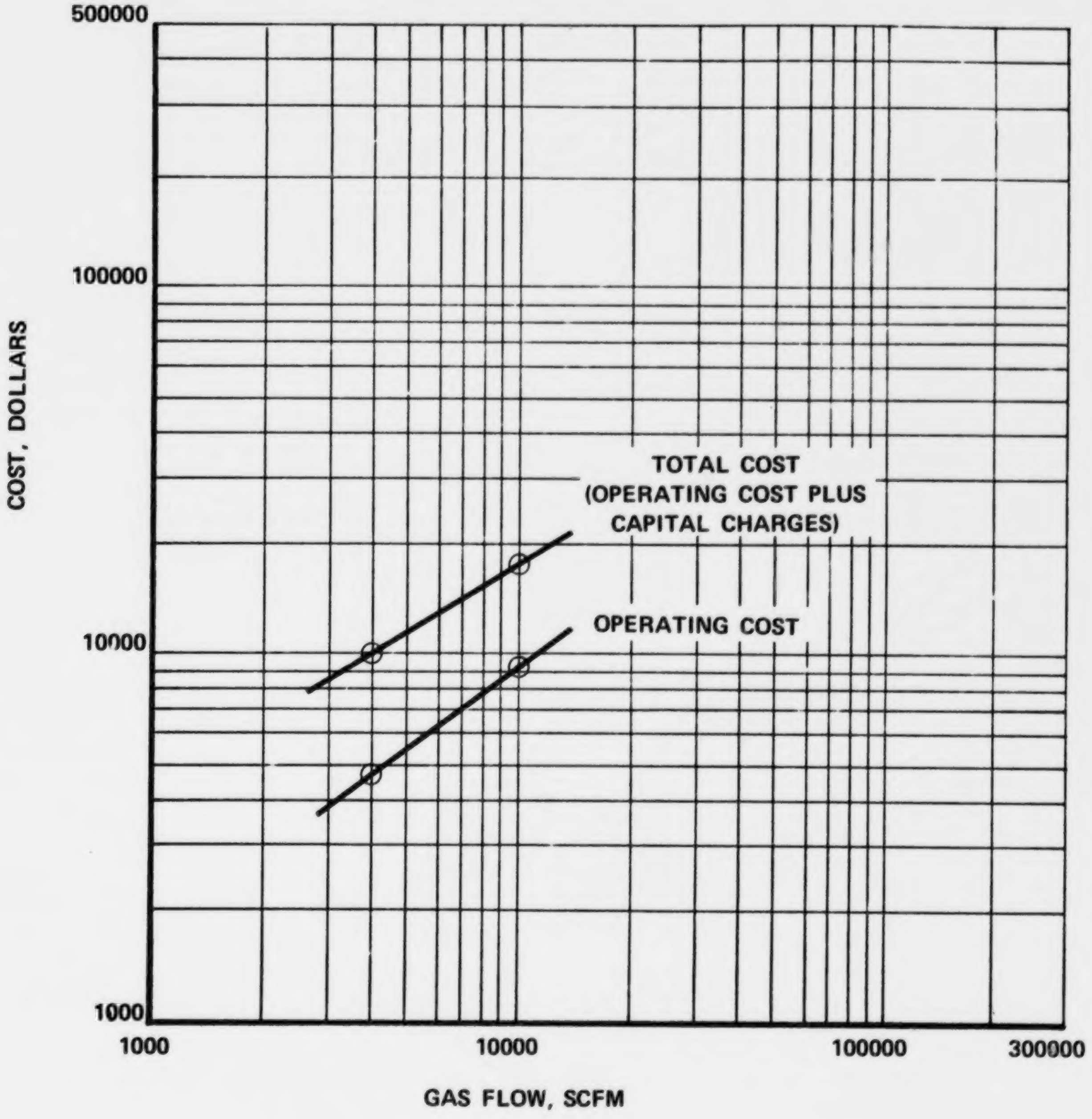


TABLE 149

THERMAL INCINERATOR PROCESS DESCRIPTION FOR GRAVURE PRINTING SPECIFICATION

This specification describes the requirements of a thermal combustion system for the abatement of hydrocarbon emissions from gravure printing facilities. A schematic flow diagram of the desired combustion system is included in Figure 115 while processing conditions and specifications for small and large facilities are tabulated in Table 150.

The incinerator will be gas fired using natural gas available at a pressure of 1.0 psig and having a specific gravity of 0.6 and an upper heating value of 1,040 Btu/SCF. The exhaust gas from the printing press will contain sufficient oxygen (greather than 16%) to allow firing of the burner without the addition of a combustion air system.

A fan equipped with a V-belt drive will be required at the incinerator inlet. The fan will have the capacity to overcome the pressure drop of the ductwork, burner, and heat exchanger. The system's ductwork will be sized for a maximum ΔP of 2 in. w.c. (hot).

The heat exchanger will be a parallel flow shell and tube exchanger and be designed to operate at an outlet incinerator temperature of 1,500°F. Maximum exchanger pressure drop, both sides, will not exceed 6.0 in. w.c. (hot) and dirty gas will flow through the tube side. The approximate thermal efficiency of the exchanger is tabulated in the specification sheet.

The incinerator will be supplied with a suitable control panel, and all equipment will be designed for outdoor operation and to meet Factory Insurance Association's standards.

The cost estimate for the combustion system will include the following items:

Incinerator
Burner
10 ft stack
Controls
Control panel
Structural steel
Fuel gas piping
Electrical
Ductwork
Insulation
Fan
Fan motor
Two day start-up service
Heat exchanger
Dampers

All items with exception of the incinerator, burner, and heat exchanger will be considered auxiliaries.

The incinerator will be located on the roof of the printing facility, and no modification of the building structure is required. All utilities are available within 30 ft of the control cabinet, motor, and burner.

Since the hydrocarbon concentration of the exhaust gas from this gravure printing operation is greater than 25% of the LEL, the exhaust gas will be diluted with sufficient hot recycle gas to lower the hydrocarbon concentration to 25% LEL.

FIGURE 115

SCHEMATIC FLOW DIAGRAM OF
THERMAL COMBUSTION WITH
HEAT EXCHANGE FOR
GRAVURE PRINTING

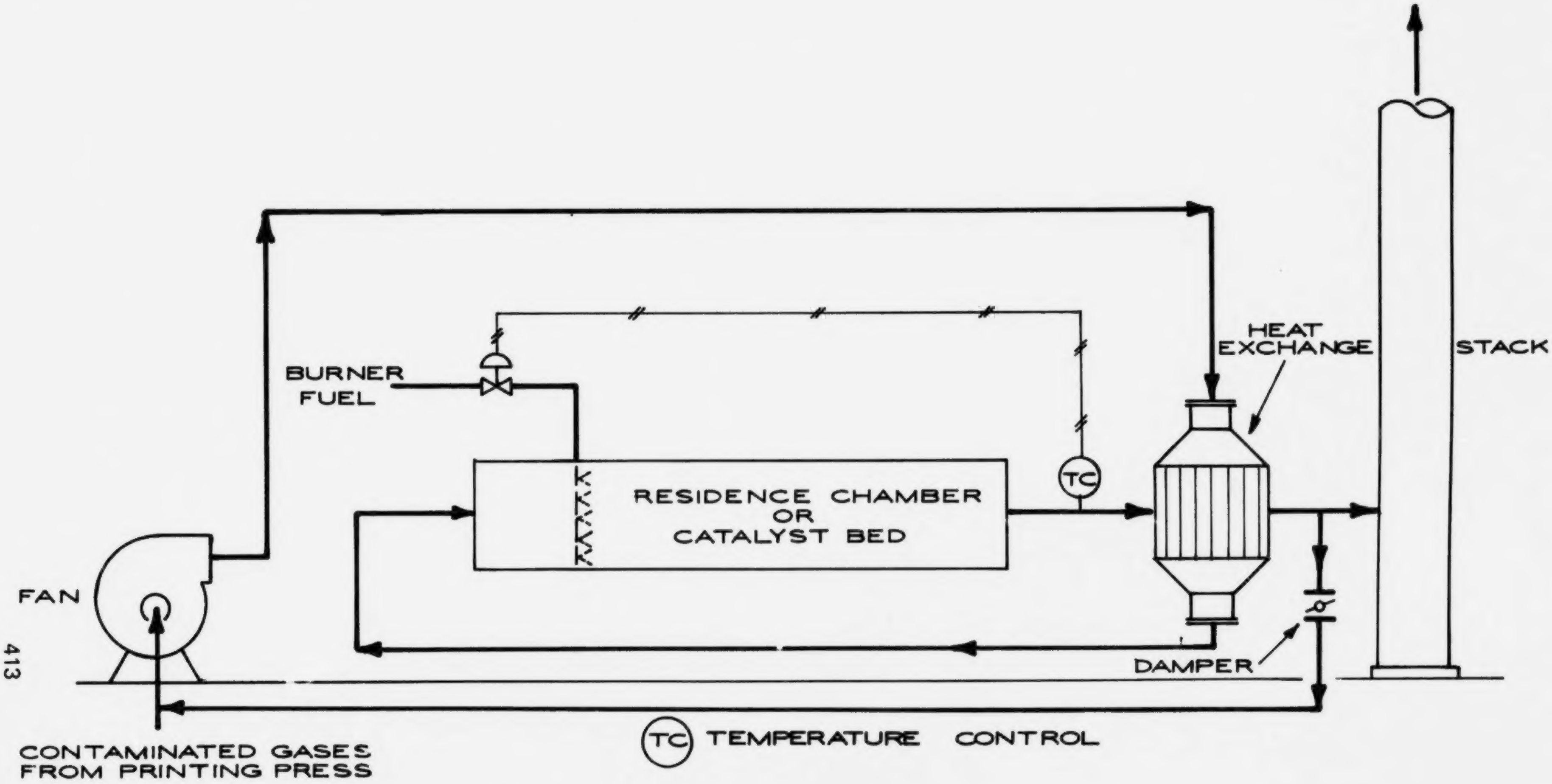

HEAT
EXCHANGE
STACK
BURNER
FUEL
TC
RESIDENCE CHAMBER
OR
CATALYST BED
FAN
DAMPER
CONTAMINATED GASES
FROM PRINTING PRESS
TC TEMPERATURE CONTROL

TABLE 150

THERMAL INCINERATOR OPERATING CONDITIONS FOR GRAVURE PRINTING SPECIFICATION (WITH HEAT EXCHANGE)

Plant Size	Small	Large
Process Conditions		
Effluent Gas Temperature	90	90
Effluent Gas Rate, SCFM	12,000	24,000
Effluent Gas Rate, ACFM	12,500	25,000
Hcbn Concentration, % LEL	35	35
Hcbn Emission Rate, lbs/hr	621	1,242
Heating Value of Gas, Btu/SCF	15.2	15.2
Heat of Combustion of Hcbn, Btu/lb(1)	17,600	17,600
Incinerator Specifications		
Inlet Hcbn Concentration, % LEL(2)	25% (max)	25% (max)
Residence Time @ Temperature, sec	0.6	0.6
Inlet Tubeside Temp., °F	90	90
Unit Outlet Temperature, °F	1,500	1,500
Burner Δ T from Fuel Gas, °F	25°F (min)	25°F (min)
Thermal Efficiency	To obtain min burner Δ T ≃ 40%	

(1) Assumed as toluene
(2) To be obtained with hot recycle gas dilution

Preceding page blank

TABLE 151

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR THERMAL INCINERATORS FOR GRAVURE PRINTING (WITH HEAT EXCHANGE)

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			12,500	25,000
°F			90	90
SCFM			12,000	24,000
Moisture Content, Vol %				
Effluent Hydrocarbon				
%LEL			35	35
lb/hr			621	1,242
Btu/SCF			15.2	15.2
Cleaned Gas Flow				
ACFM			44,750	89,500
°F			1,500	1,500
SCFM			12,100	24,200
Moisture Content, Vol. %			-	-
Cleaned Gas Contaminant Loading				
lb/hr			<31	<62
Cleaning Efficiency, %			>95	>95
(1) Gas Cleaning Device Cost			47,600	72,350
(2) Auxiliaries Cost				
(a) Fan(s)				
(b) Pump(s)			5,125	10,578
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			3,100	3,750
(b) Foundations & Support			3,750	5,850
Ductwork			3,245	4,870
Stack			950	1,600
Electrical			1,000	1,654
Piping			275	488
Insulation				
Painting				
Supervision			1,050	1,200
Startup			450	550
Performance Test			1,750	1,750
Other			5,960	6,260
(4) Total Cost			74,255	110,900

FIGURE 116

CAPITAL COSTS FOR THERMAL INCINERATORS FOR GRAVURE PRINTING (WITH HEAT EXCHANGE)

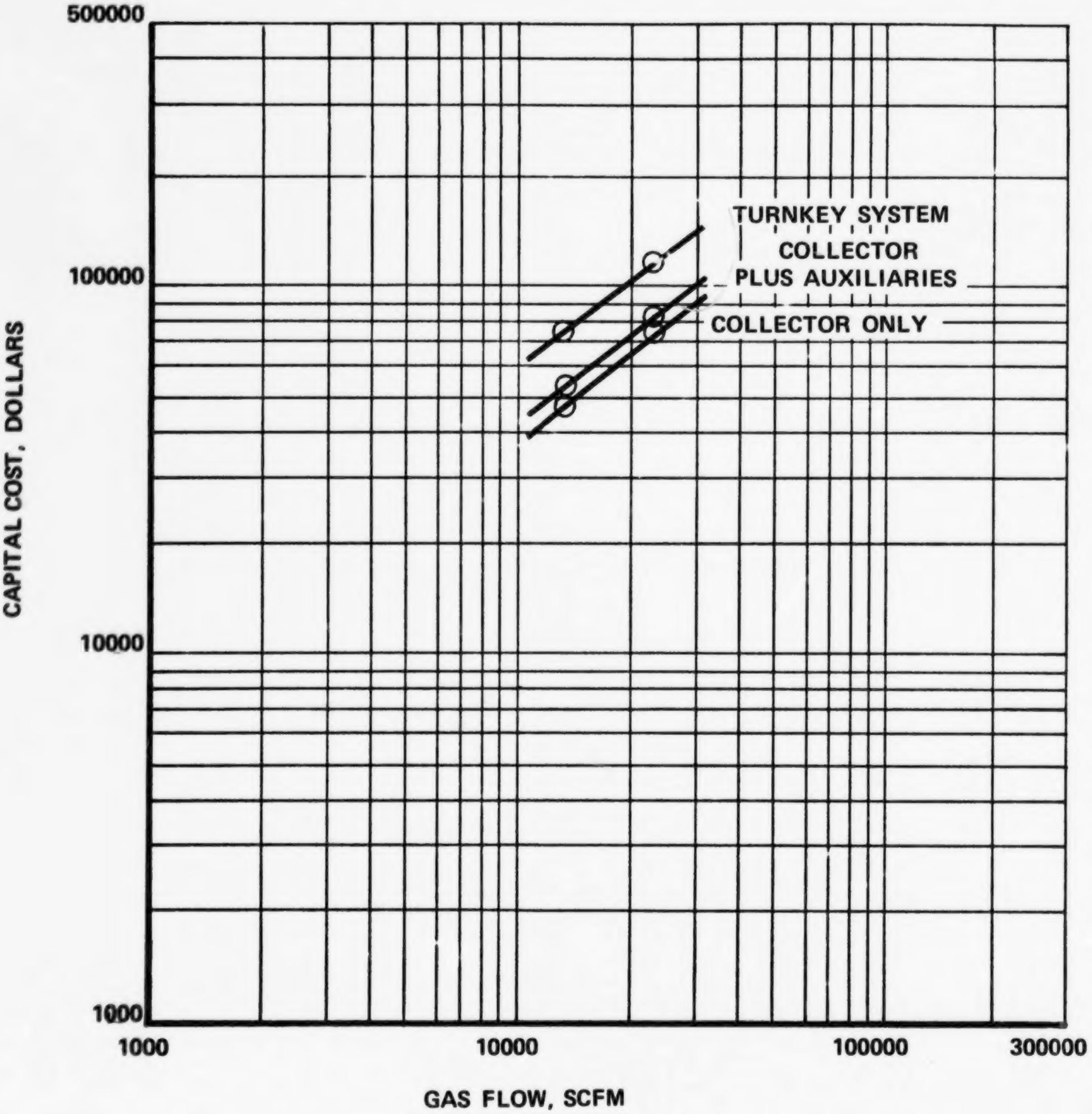


TABLE 152

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR THERMAL INCINCERATORS FOR GRAVURE PRINTING (WITH HEAT EXCHANGE)

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				4,000	4,000
Operating Labor (if any)					
Operator	$6/hr			1,575	1,575
Supervisor					
Total Operating Labor				1,575	1,575
Maintenance					
Labor	$6/hr			360	360
Materials				135	260
Total Maintenance				495	620
Replacement Parts				125	125
Total Replacement Parts				125	125
Utilities					
Electric Power				1,796	4,334
Fuel				6,520	13,020
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				8,316	17,354
Total Direct Cost				10,511	19,674
Annualized Capital Charges				7,426	11,090
Total Annual Cost				17,937	30,764

FIGURE 117

ANNUAL COSTS FOR THERMAL INCINERATORS
FOR GRAVURE PRINTING
(WITH HEAT EXCHANGE)

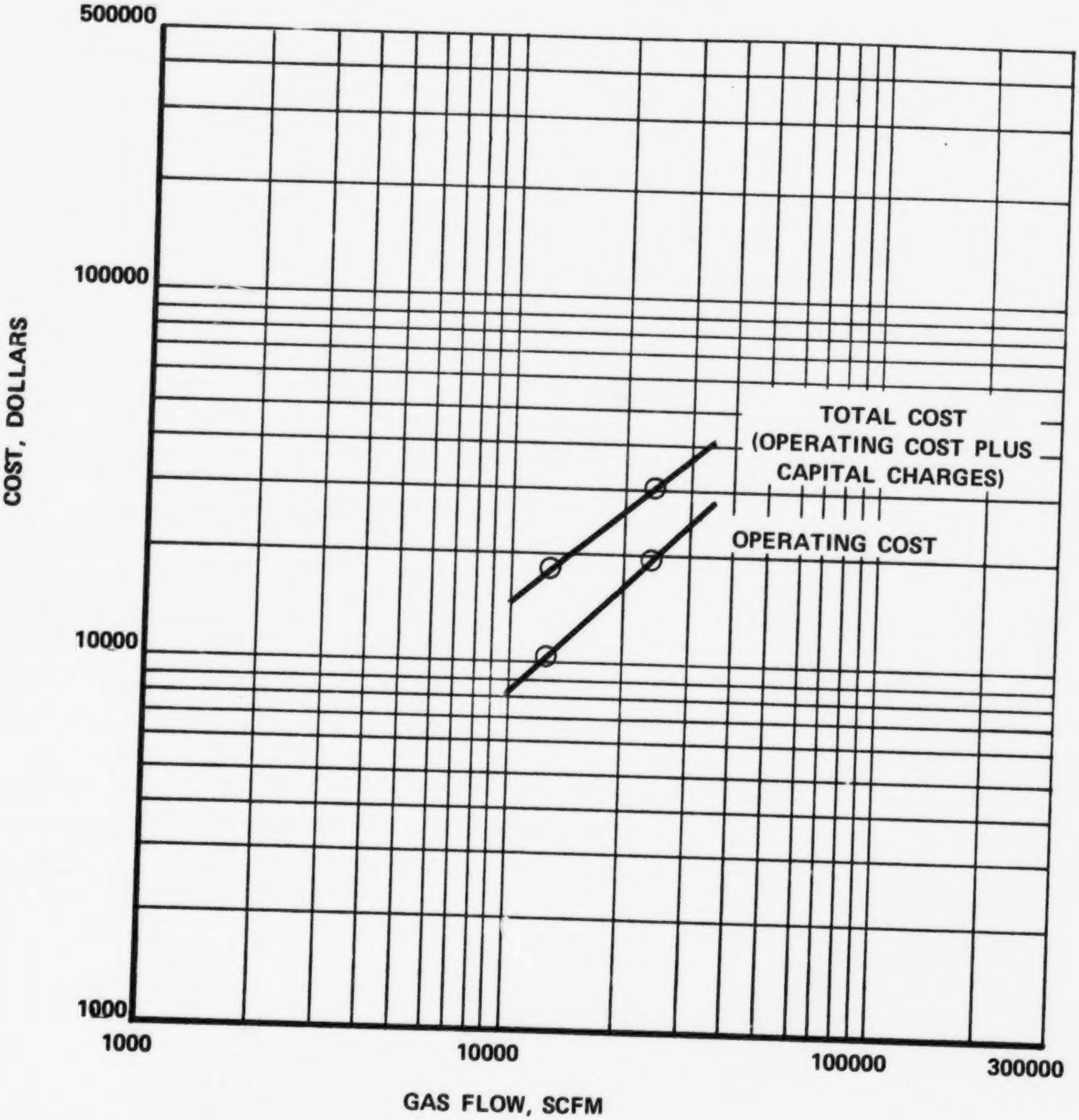


TABLE 153

CARBON ADSORPTION PROCESS DESCRIPTION FOR GRAVURE PRINTING SPECIFICATION

A carbon adsorption system is to remove hydrocarbon emissions from a gravure printing facility and to provide for their recovery for reuse at the facility. Effluent gas from each of the presses will be combined and treated in the single adsorption system. Process conditions and specifications for a small and large facility are tabulated in Table 154.

The system will be of the continuous, regenerative type. Regeneration will be accomplished with low pressure steam and air. The hydrocarbon to be recovered will be toluene. Suitable carbon will be selected for this sytem. The system will be designed for automatic operation but will have manual overrides. In order to insure adequate ventilation, the ducts leading from each of the presses will be designed for equal pressure losses. Bids will include the following:

1. *Carbon adsorption devices.*
2. *A booster fan sufficient to handle the total gas effluent from the vent fans over each press and to overcome the pressure drop across the system's ductwork and across the carbon adsorption beds.*
3. *A continuous type filter located before the carbon beds. The filter will remove ink and dust particles, in order to prevent clogging and/or fouling of the carbon beds.*
4. *Two hundred fifty feet of rectangular ductwork which will provide for the collection of the effluent gas from the different presses and provide for their transport after their exit from the adsorption system. The manifold duct will be left open at one end in order to provide for capacity differences due to changes in press operation.*
5. *An automatic control system which continually measures and records the toluene level at the outlet of the adsorption system.*
6. *Appropriate recovery and storage facilities for the hydrocarbon solvent.*
7. *Dampers.*

All items other than the first will be considered auxiliaries.

The adsorption system will be located on the roof and no modifications of the existing structure will be necessary. Low pressure steam, city water at 90°F, and electric power at 440 v, 220 v, and 110 v are all available within 30 ft of the site.

TABLE 154

CARBON ADSORPTION OPERATING CONDITIONS FOR GRAVURE PRINTING SPECIFICATION

	Small	*Large*
Number of Presses	*4*	*12*
ACFM/Press	*12,450*	*12,450*
Adsorption Equipment Inlet		
Gas Rate, ACFM	*49,800*	*149,400*
Temperature, °F	*90*	*90*
Gas Rate, SCFM	*48,000*	*144,000*
Hydrocarbon Solvent		
% LEL	*35*	*35*
ppm by vol	*3,560*	*3,560*
lb/hr	*2,484*	*7,452*
Relative Humidity, %	*≤ 50*	*≤ 50*
Heating Value, Btu/SCF	*15.2*	*15.2*
Performance Requirement		
Hydrocarbon Recovery, %	*≥ 95*	*≥ 95*

TABLE 155

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR CARBON ADSORPTION FOR GRAVURE PRINTING

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			49,800	149,400
°F			90	90
SCFM			48,000	144,000
Moisture Content, R.H.%			50	50
Effluent Contaminant Loading				
			3,560	3,560
lb/hr			2,484	7,452
Cleaned Gas Flow				
ACFM			50,990	152,970
°F			105	105
SCFM			47,836	143,508
Moisture Content, Vol. %				
Cleaned Gas Contaminant Loading				
ppm			178	178
lb/hr			124	372
Cleaning Efficiency, %			95	95
(1) Gas Cleaning Device Cost			215,150	504,516
(2) Auxiliaries Cost				
*(a) Fan(s)			9,850	20,950
*(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			172,567	325,867
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other - Carbon			25,029	61,901
(4) Total Cost			422,596	913,234

*Included in item (1)

FIGURE 118

CAPITAL COSTS FOR CARBON ADSORPTION FOR GRAVURE PRINTING

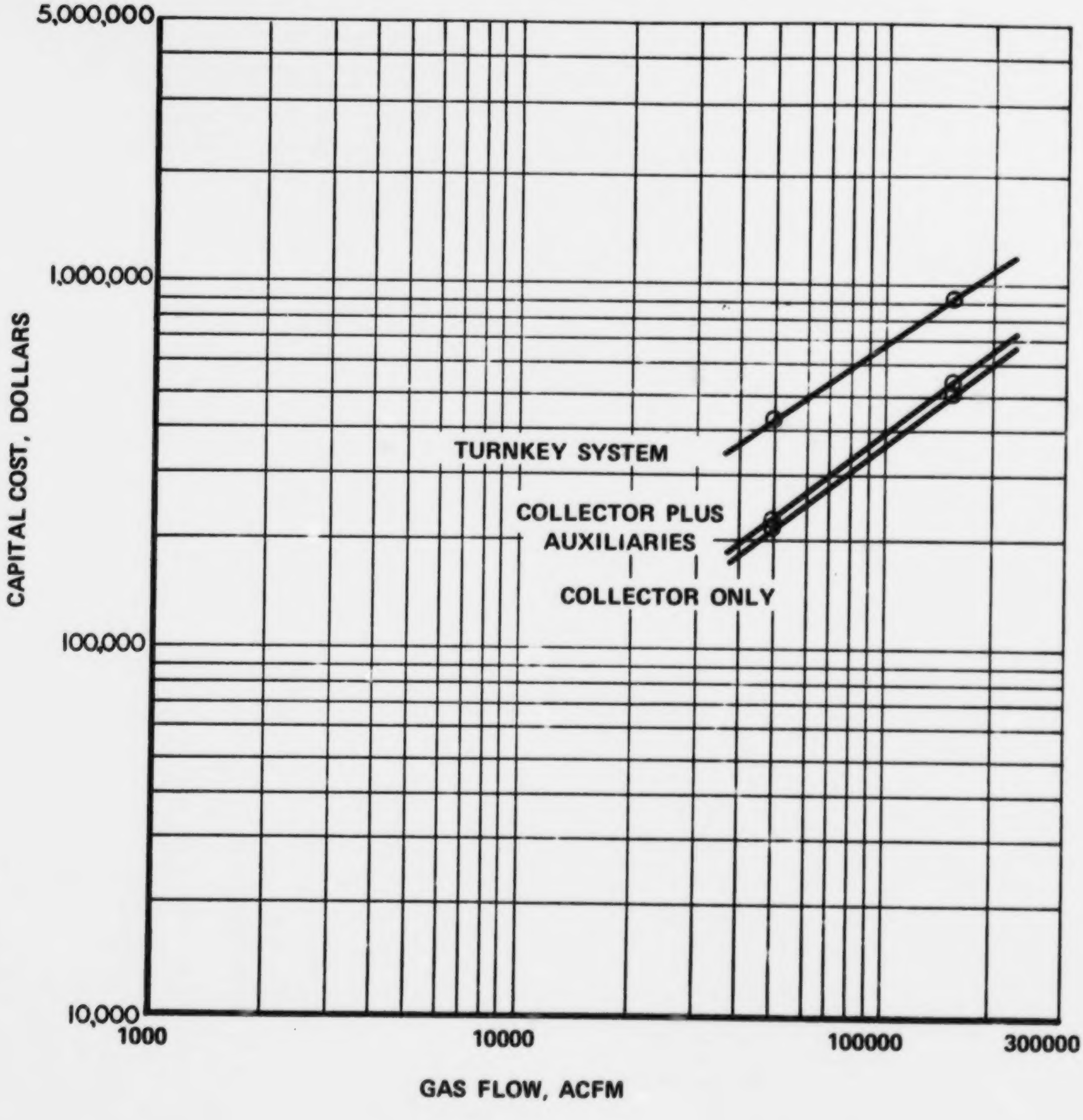


TABLE 156

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR CARBON ADSORPTION FOR GRAVURE PRINTING

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)				8,000	8,000
Operator					
Supervisor	$6/hr				
Total Operating Labor				2,844	2,844
Maintenance				2,844	2,844
Labor					
Materials					
Total Maintenance					
Replacement Parts				4,973	10,967
Total Replacement Parts					
Utilities				3,346	7,427
Electric Power					
Fuel	$.011/kw-hr				
Water (Process)				12,833	38,148
Water (Cooling)	$.05/M Gal				
Steam (low pressure)	$1.00/M lb			14,975	44,925
Carbon make-up 4%/yr				74,667	210,667
Total Utilities				1,001	2,476
Total Direct Cost				103,476	296,216
Annualized Capital Charges				114,639	317,454
Total Annual Cost				42,260	91,324
				156,899	408,778

FIGURE 119

ANNUAL COSTS FOR CARBON ADSORPTION FOR GRAVURE PRINTING

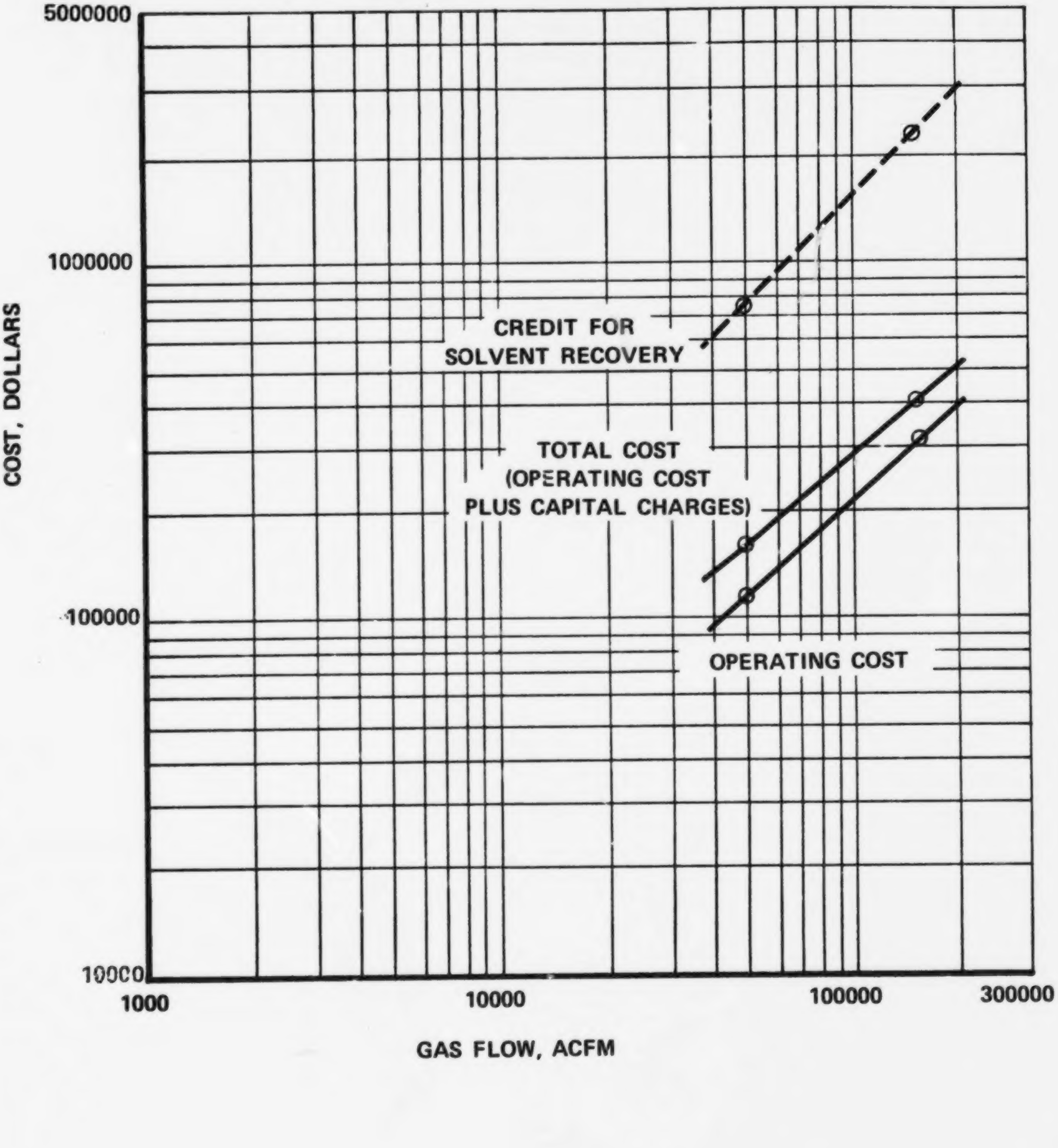


TABLE 157

CONFIDENCE LIMITS FOR CAPITAL COST OF CARBON ADSORPTION FOR GRAVURE PRINTING

Population Size – 20

Sample Size – 3

Capital Cost = $422,629

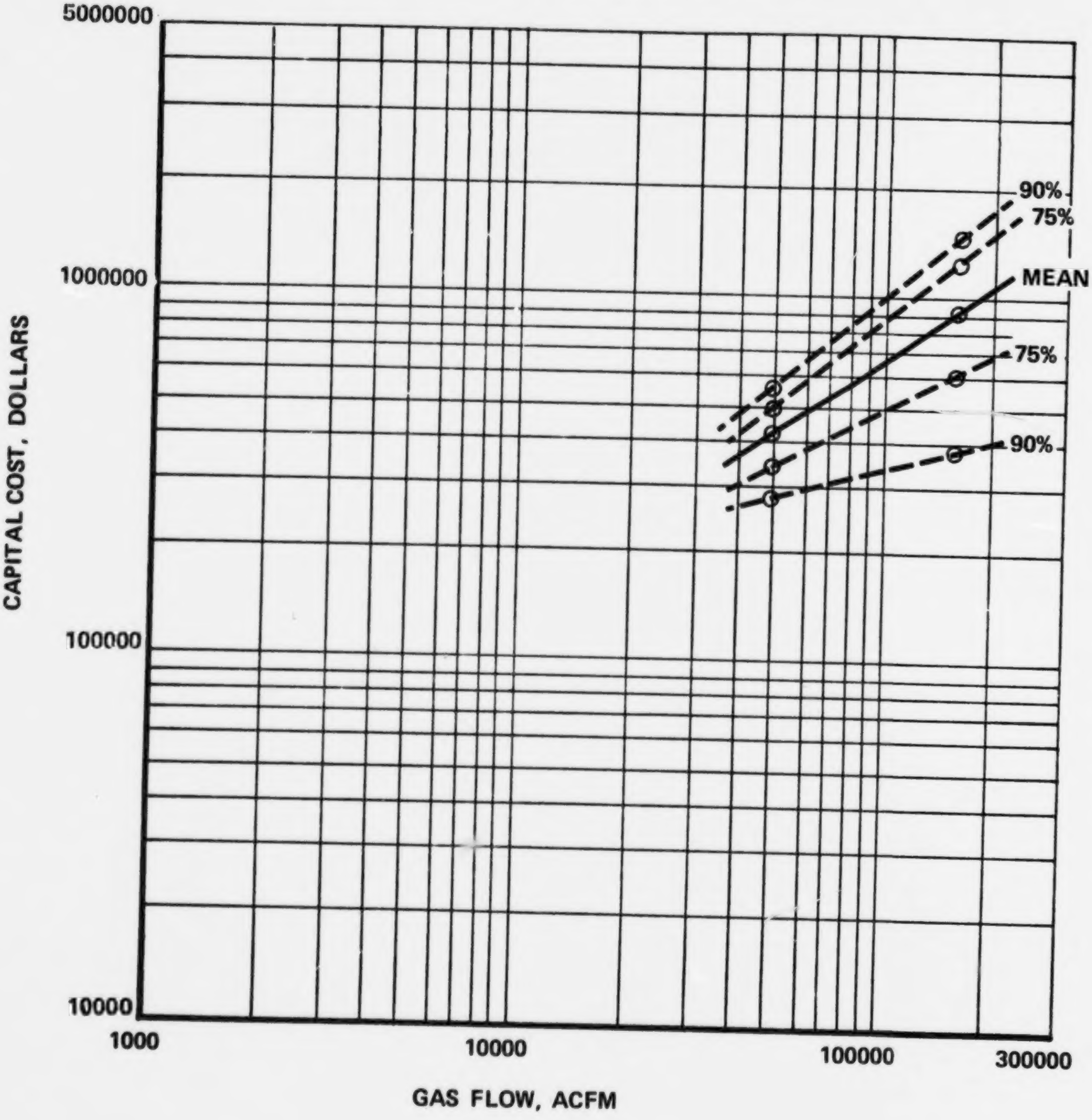
Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$382,846	$462,412
75	345,547	499,711
90	288,152	557,106
95	235,079	610,179

Capital Cost = $913,235

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$753,573	$1,072,900
75	603,877	1,222,590
90	373,531	1,452,940
95	160,533	1,665,940

FIGURE 120

CONFIDENCE LIMITS FOR CAPITAL COST OF CARBON ADSORPTION FOR GRAVURE PRINTING



REFERENCES

1. U.S. Department of Commerce, *U.S. Industrial Outlook 1970*, Washington, Government Printing Office, 1969, Chapter 5.

2. Gadomski, R. R., David, M. P., Blahut, J. A., *Evaluation of Emissions and Control Technologies in the Graphic Arts Industries*, Graphic Arts Technical Foundation, 1970.

3. U.S. Department of Commerce, *1967 Census of Manufacturers Preliminary Reports, Industry Series*, Washington, U.S. Department of Commerce, 1969.

4. Anon., *Printing Plates Magazine*, Vol. 56, No. 4, April, 1970, p. 28.

5. Drury, H. F., *Total Products Shipments – 1970, Inland Printer/American Lithographer*, Vol. 164, No. 4, Jan., 1970, p. 46-48.

6. Anon., *Survey of Gravure Printing*, Gravure Technical Association Bulletin, Vol. 20, No. 2 Summer, 1969, p. 10-24.

7. Cartright, H. M., Jackay, R., *Rotogravure – A Survey of European and American Methods*, Mackay Publishing Co., 196, p. 240.

8. Hays, D. R., *Official Digest*, Vol. 36, No. 473, 1963, pp. 605-624.

9. Bock, R. F., *Ink Drying – GATF Research Department Report of Progress 1969*, Pittsburgh, Pa., Graphic Arts Technical Foundation, 1970.

10. Heuser, A. R., *Friends in Air Conservation in the Graphic Arts Industry*, Gravure, Vol. 15, No. 2, Feb., 1969, pp. 22-25, 45-48.

5. SOAP AND DETERGENTS

Soaps and synthetic detergents are two of the most common products in the Western World. With approximately 14.5 million tons of soap and detergent produced in 1969,(1) they may be called the most widely used chemical compounds in the modern world. U. S. production of synthetic detergents (syndets) in 1968 was reported to be 2.75 million tons; this was 83% of total soap and detergent consumption which was approximately 3.3 million tons.(1)

Water, which wets unevenly, is not a very efficient cleaning compound. The major function of all soaps and detergents is to reduce the surface tension and improve the wetting efficiency of the water. The surface tension prevents the water from fully contacting the material to be cleaned and removing the dirt. Soaps and detergents have both hydrophilic and hydrophobic groups. The hydrophilic group enables the detergent to dissolve in water and reduce surface tension, while the hydrophobic part contacts the dirt. The dirt is removed by agitation and prevented from resettling by electrical charges produced by the detergent.(2)

Soap making is an ancient practice, dating back to Roman civilization. However, the first synthetic detergent was not made until World War I. Around 1930, long chain alkyl-aryl sulfonates were introduced in the U. S. and began to compete with soap. In the early 1950's, tetra-propylene benzene sulfonate (TPBS) dominated the synthetic detergent market. This detergent caused extensive foam formation in inland waterways. TPBS has a branched alkyl chain and, consequently, resists biological degradation. In 1965, U. S. detergent manufacturers switched to linear alkyl sulfonates (LAS), a biodegradable product with detergency values comparable with TPBS. Recently, the use of sodium tripolyphosphate as a builder in most detergent formulations has been suspected of encouraging algae growth in natural waterways, so additional formulation changes may be expected in the near future.

PROCESS DESCRIPTION

This section will deal mainly with synthetic detergent manufacture. The major source of air pollution in detergent production is spray drying. The air pollution problems found in soap making are generally not as severe as those in detergent making because soap flakes are primarily produced in a drum dryer. Only one-third of the soap produced is spray dried to flake or powder form.(1)

Modern detergent formulations are extremely varied. The active ingredient, usually linear alkyl benzene sulfonate, alcohol sulfate, alcohol ether sulfonate, alcohol ethoxylate, or sodium alkenyl sulfonate, comprises a third or less of the detergent weight. The remainder of the detergent consists of a sequestering agent, an anti-corrosion agent, an anti-soil redeposition agent, a foam booster, bleach, a brightener, filler, and in some cases, enzymes.(3)

Basically, the manufacture of powdered detergents consists of the following steps:

1. Sulfation or sulfonation of the raw material (fatty alcohol or linear alkylate).
2. Neutralization of the acid to produce the sulfate or sulfonate.
3. Blending the sulfonate paste with other ingredients.
4. Spray drying of this mixture.
5. Packaging and storing of the detergent.

In the manufacture of liquid detergents, a foam stabilizer and solubilizer is blended in before the detergent is bottled. Liquid detergents accounted for 20 to 25% of total syndet production in 1968.

Detergents and soaps may be broken down into three main classes:

1. Anionic detergents – ionize in water to produce a negatively charged organic ion; used mainly in soaps and powdered synthetic detergents.
2. Nonionic detergents – do not ionize in water; used mainly in liquid synthetic detergents.
3. Cationic detergents – ionize in water to produce a positively charged organic ion; not normally classified as detergents, but as germicides.

In 1966, anionic detergents (including soap) represented 74.3% of total detergent production, nonionic, 20.6%, and cationic, less than 5%.(4)

The most frequently used active ingredient in anionic synthetic detergents today is linear alkylbenzene sulfonate (LAS). Alcohol sulfates and alpha-olefin sulfonates (AOS) are also used. Although AOS consumption in 1968 was about 19% of total sulfonate consumption, it has been discussed as the most serious competition to LAS.(2) Overall routes to these three ingredients are shown in Figure 121.(2,5) The major feed materials to the detergent manufacturing plant for these three processes would be linear alkylbenzene, primary alcohol, and linear paraffin wax (or ethylene), respectively.

Of the nonionic active ingredients in synthetic detergents, ethoxylated long chain alcohols and to a lesser extent, ethoxylated alkyl-phenols are the most common. These materials are produced by reacting the linear alcohol or alkyl-phenol with ethylene oxide in a mole ratio of 1:(2 to 10).(6)

The sulfonation (or sulfation) of the raw materials in anionic synthetic detergent production is very similar for the three main processes. Flow sheets are shown in Figures 122 and 123 for the sulfonationof linear alkyl benzene and linear alpha-olefins.(2) In the actual sulfonation reaction, a mixture of 4% SO_3 (vol/vol air) is generally used. For optimum reaction conditions, this mixture should be at a pressure of 1.0 to 1.3 atmospheres, a temperature of 100 to 140° F, a dew point of −58° F, and free of acid mists.(2) This SO_3/air mixture is usually prepared from one of three reagents. Stabilized liquid sulfur trioxide is probably the most versatile. It is exclusively used in alpha-olefin sulfonation. Sulfuric acid, usually in the form of oleum ($H_2SO_4 \cdot SO_3$), and occasionally sulfur are also used to produce the SO_3/air mixture.

In some plants oleum or less frequently sulfuric acid is directly mixed with alkyl benzene to produce the sulfonate. However, large excesses of oleum and larger excesses of sulfuric acid are required to drive the reaction to completion, causing these processes to be relatively expensive.(7)

Sulfonation processes are typically composed of a raw material metering system, a sulfonation reactor, a cooling circuit, a separating circuit, and a hydrolyzer.

In the manufacture of linear alkyl benzene sulfonate, the linear alkyl benzene and SO_3/air mixture are metered into the reactor at a SO_3/raw material mole ratio of 1.0 to 1.1.(2) The reaction is very quick and exothermic, and heat must be rapidly removed to prevent discoloration. The heat of reaction is removed by recycling the reactants through a temperature controlled heat exchanger. The reaction is completed as the overflow from the reactor flows through a reaction coil. Although feed materials other than linear alkyl benzene (alpha-olefins, fatty alcohols) may require different post-treatment, the sulfonation step is much the same.

Typical operating conditions for the sulfonation of linear benzene

FIGURE 121

OVER-ALL ROUTES TO LINEAR ALKYLATE SULFONATES (LAS), ALPHA-OLEFIN SULFONATES (AOS) AND ALCOHOL SULFATES

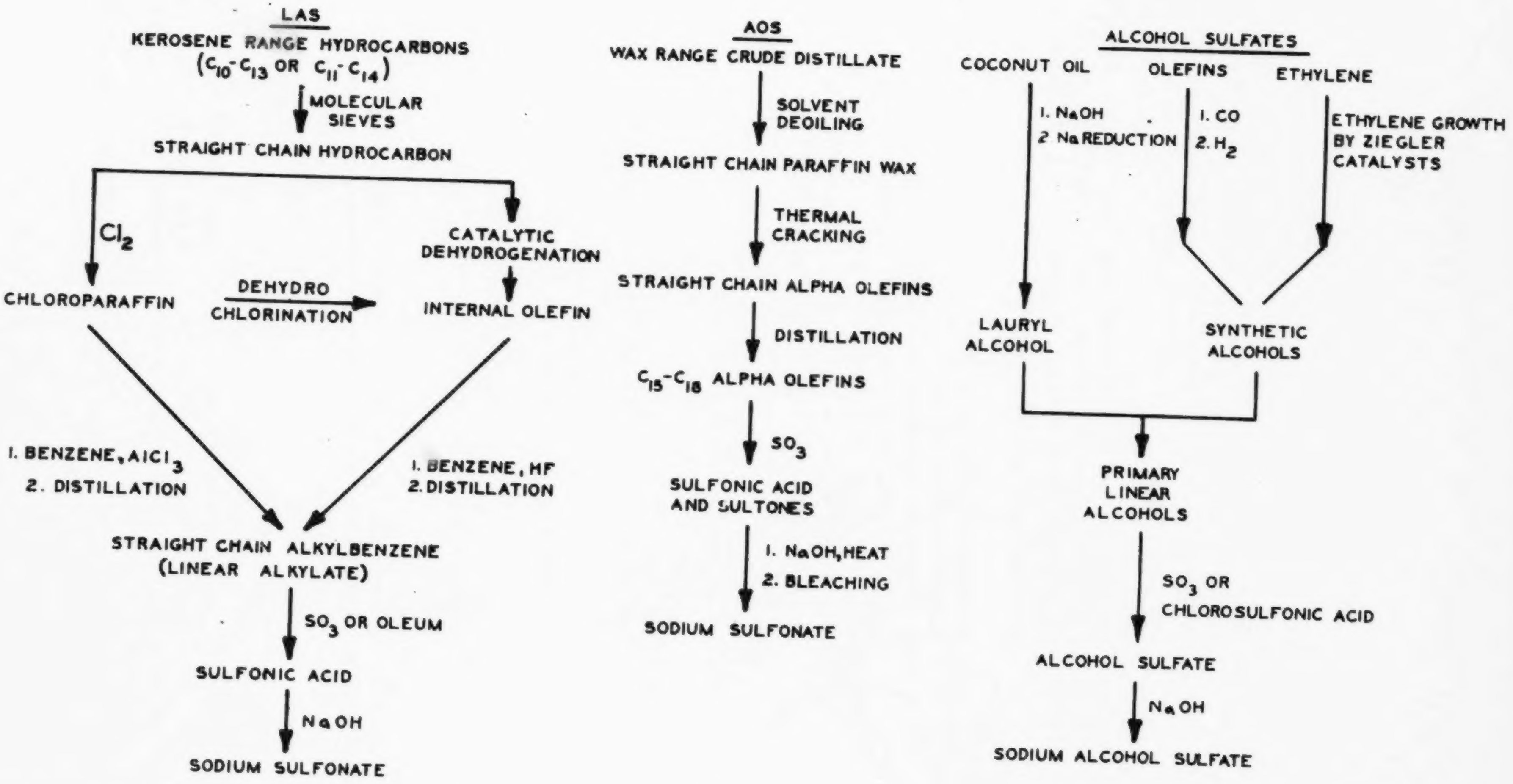

LAS
KEROSENE RANGE HYDROCARBONS
(C_{10}-C_{13} OR C_{11}-C_{14})
MOLECULAR SIEVES
STRAIGHT CHAIN HYDROCARBON
Cl_2
CATALYTIC DEHYDROGENATION
CHLOROPARAFFIN
DEHYDRO CHLORINATION
INTERNAL OLEFIN
1. BENZENE, $AlCl_3$
2. DISTILLATION
1. BENZENE, HF
2. DISTILLATION
STRAIGHT CHAIN ALKYLBENZENE (LINEAR ALKYLATE)
SO_3 OR OLEUM
SULFONIC ACID
NaOH
SODIUM SULFONATE
AOS
WAX RANGE CRUDE DISTILLATE
SOLVENT DEOILING
STRAIGHT CHAIN PARAFFIN WAX
THERMAL CRACKING
STRAIGHT CHAIN ALPHA OLEFINS
DISTILLATION
C_{15}-C_{18} ALPHA OLEFINS
SO_3
SULFONIC ACID AND SULTONES
1. NaOH, HEAT
2. BLEACHING
SODIUM SULFONATE
ALCOHOL SULFATES
COCONUT OIL
OLEFINS
ETHYLENE
1. NaOH
2. Na REDUCTION
1. CO
2. H_2
ETHYLENE GROWTH BY ZIEGLER CATALYSTS
LAURYL ALCOHOL
SYNTHETIC ALCOHOLS
PRIMARY LINEAR ALCOHOLS
SO_3 OR CHLOROSULFONIC ACID
ALCOHOL SULFATE
NaOH
SODIUM ALCOHOL SULFATE

FIGURE 122

LOW- AND HIGH-ACTIVE SULFONATION USING OLEUM

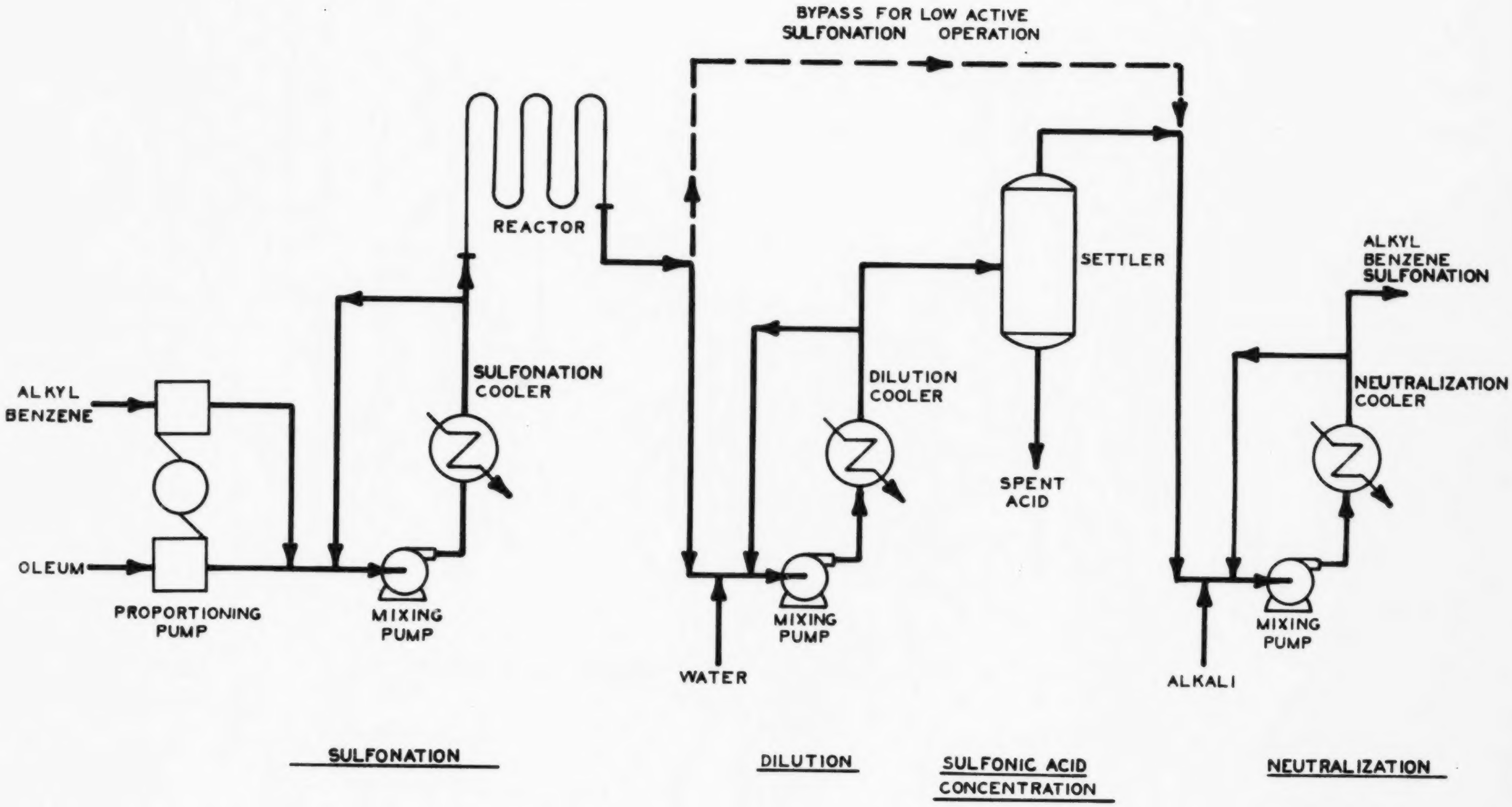
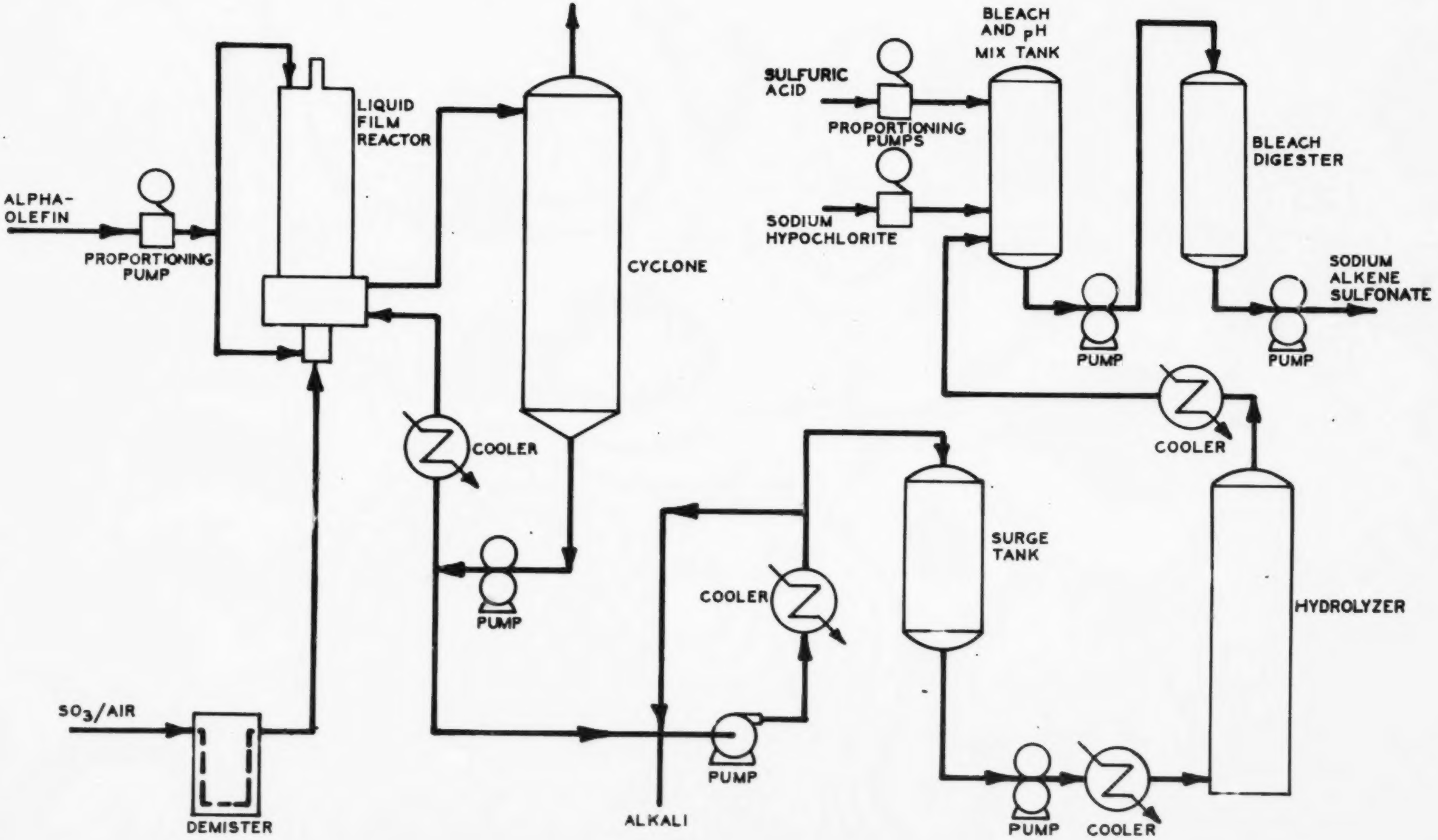

BYPASS FOR LOW ACTIVE SULFONATION OPERATION
REACTOR
SETTLER
ALKYL BENZENE SULFONATION
ALKYL BENZENE
OLEUM
PROPORTIONING PUMP
SULFONATION COOLER
MIXING PUMP
DILUTION COOLER
MIXING PUMP
WATER
SPENT ACID
NEUTRALIZATION COOLER
MIXING PUMP
ALKALI
SULFONATION
DILUTION
SULFONIC ACID CONCENTRATION
NEUTRALIZATION

FIGURE 123

ALPHA - OLEFIN SULFONATION FLOWSHEET


ALPHA-OLEFIN
PROPORTIONING PUMP
LIQUID FILM REACTOR
CYCLONE
COOLER
PUMP
SO$_3$/AIR
DEMISTER
ALKALI
PUMP
COOLER
SURGE TANK
PUMP
COOLER
HYDROLYZER
COOLER
SULFURIC ACID
PROPORTIONING PUMPS
SODIUM HYPOCHLORITE
BLEACH AND pH MIX TANK
PUMP
BLEACH DIGESTER
PUMP
SODIUM ALKENE SULFONATE

alkylates are given in Table 158.[7] Reactor capacities range from less than 1 million lb/yr to more than 20 million lb/yr of sulfonic acid, based on 6000 hr/yr of operation.[7]

The reactor effluent consists mainly of sulfonic and sulfuric acids. The excess sulfuric acid is usually removed by first diluting it with water; after dilution, the difference in densities is large enough to permit phase separation and the spent acid is removed after a residence time of about 15 minutes.

The sulfonic acid is next neutralized to form the sulfonate. Caustic soda is generally used. The neutralization reaction is also rapid and extremely exothermic and the reactants are recirculated through a cooler. The solids content of this slurry will normally be about 46%.[2]

In the sulfation of primary alcohols, chlorosulfonic acid is usually employed as the sulfating agent.[8] Gaseous HCl is liberated during sulfation and this necessitates absorption and disposal of the acid, which is extremely corrosive. For this reason the reaction is usually carried out batchwise in glass-lined kettles at about 86° F.[9] Sulfuric acid can be used in alcohol sulfation but because of the reverse hydrolysis reaction, equilibrium conversion to the sulfonic acid is only 65%, and a large excess of acid must be used. It appears that SO_3 gas may be the best method for sulfation, as well as for sulfonation. An excess of 3.5% SO_3 (converter gas) provides good sulfation of linear alcohols (lauryl or synthetic), with a sulfation time of 40 minutes and mean temperature of 95° F.[8]

After sulfonation (or sulfation), the sulfonate paste is blended with the numerous detergent additives. The dry ingredients are weighed and fed to a screw conveyor which acts also as a mixer. The conveyor transports the dry ingredients to the main mixer, where they are blended with the liquid ingredients (including the sulfonate paste). The blended materials are then conveyed to an aging crutcher, where the slurry is aged and further mixed. Final drying is done in a spray dryer.

Synthetic detergent and soap powders are formed by spray drying. Since World War II, vacuum spray dryers have also been used to perform the functions of hot-air dryers in soap drying.[12] A flowsheet for a typical detergent spray drying plant is shown in Figure 124.[2]

Neat soap is generally about 60 to 70% solids, while the detergent sulfonate paste is usually 55 to 60% solids. The paste comes from the blending unit at a temperature of 120 to 175° F, is filtered, and pumped at pressures up

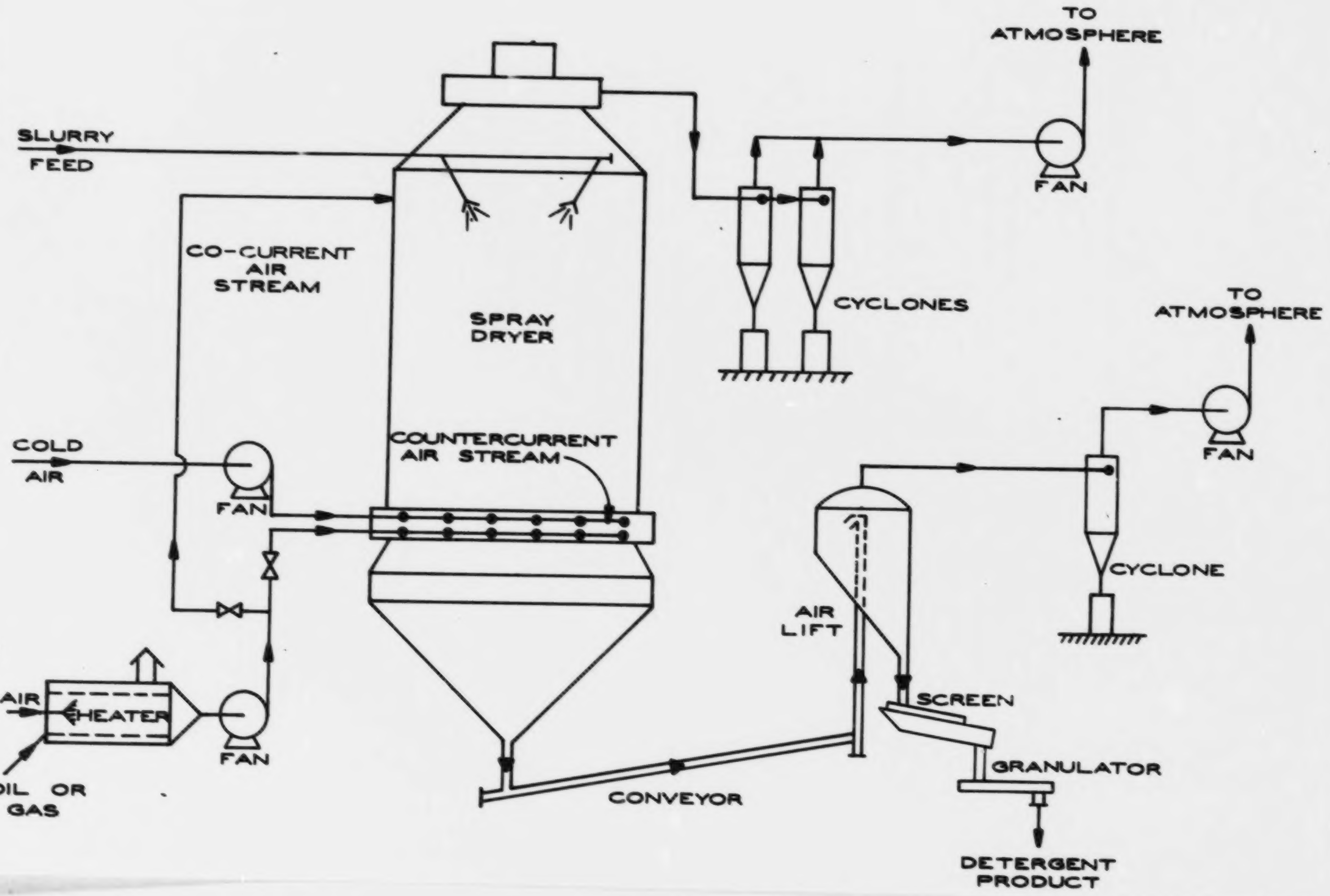
TABLE 158

TYPICAL OPERATING CONDITIONS FOR CONTINUOUS SULFONATING OF LINEAR ALKYLATE[7]

Temperature, ° F	
Linear alkylate feed	65-85
Vaporized SO_3 – dry air mixture	100-110
Product sulfonic acid	120-140
Flow rates, lb/hr	
Linear alkylate feed	1240
Vaporized SO_3	433
Dry air feed (SCFM)	800
Product sulfonic acid	1670
Molar ratio, SO_3:alkylate	1.03:1.0
SO_3 carried by air, vol %	4
Aging time, minutes	15

FIGURE 124

TYPICAL SPRAY DRYING PLANT


SLURRY FEED
CO-CURRENT AIR STREAM
COLD AIR
FAN
AIR
HEATER
OIL OR GAS
FAN
SPRAY DRYER
COUNTERCURRENT AIR STREAM
CONVEYOR
CYCLONES
TO ATMOSPHERE
FAN
AIR LIFT
SCREEN
GRANULATOR
DETERGENT PRODUCT
CYCLONE
TO ATMOSPHERE
FAN

to 70 atmg through feed nozzles at the top of the spray tower.(2) The towers are usually larger than 14 ft in diameter and about 60 to 100 ft high, with a 60° bottom cone. The nozzles spray the paste into a hot drying air stream. When the paste reaches the right density, it settles to the bottom of the dryer and is removed by a belt conveyor to packaging. Moisture contents of the dry powder bead are in the 3 to 18 wt.% range.(2) The bead is rather coarse, spherical, stable, and has a large surface area.

The air for drying is taken from the atmosphere and heated in an oil or gas fired furnace to 645 to 710°F. Three different air systems are generally employed in spray dryers:

1. Parallel or co-current
2. Countercurrent
3. Mixed air flow

In the co-current system, hot air is introduced at the top of the tower and flows down the tower with the sprayed detergent particles. In the discharge cone, the drying air is separated from the beads and drawn through a cyclone. Cool air, drawn in at the bottom of the tower, cools the beads and separates the hollow beads from the fines, which are collected in the cyclone. The product produced in this system will have density of 0.08 to 0.15 g/cc and a moisture content of 3 to 8%.(2) The countercurrent system is most often used in spray drying. Hot air enters at the bottom of the tower and contacts the downward flowing detergent beads. An average tower will dry 10,000 ton/yr with larger towers handling up to 100,000 ton/yr.(3) The product bead will have a 0.18 to 0.4 g/cc density and a 6 to 15% moisture content.(2) In the mixed air flow system a reduced amount of hot air enters through the top, while a large quantity of cold air enters at the bottom. The upper part of the tower consequently operates as a co-current unit, while the lower part acts as a countercurrent unit. The powder produced is in the form of a hollow bead containing a very high percentage of water of crystallization; the water content is usually 20 to 25%.

The detergent beads are conveyed from the bottom of the drying tower to a pneumatic air lift. The beads are cooled as they travel up the air lift. At the top of the lift they are separated from the air stream by a cyclone or gravity separator and fall to a vibrating screen which removes the oversized particles. The detergent is then sprayed with perfume and packaged.

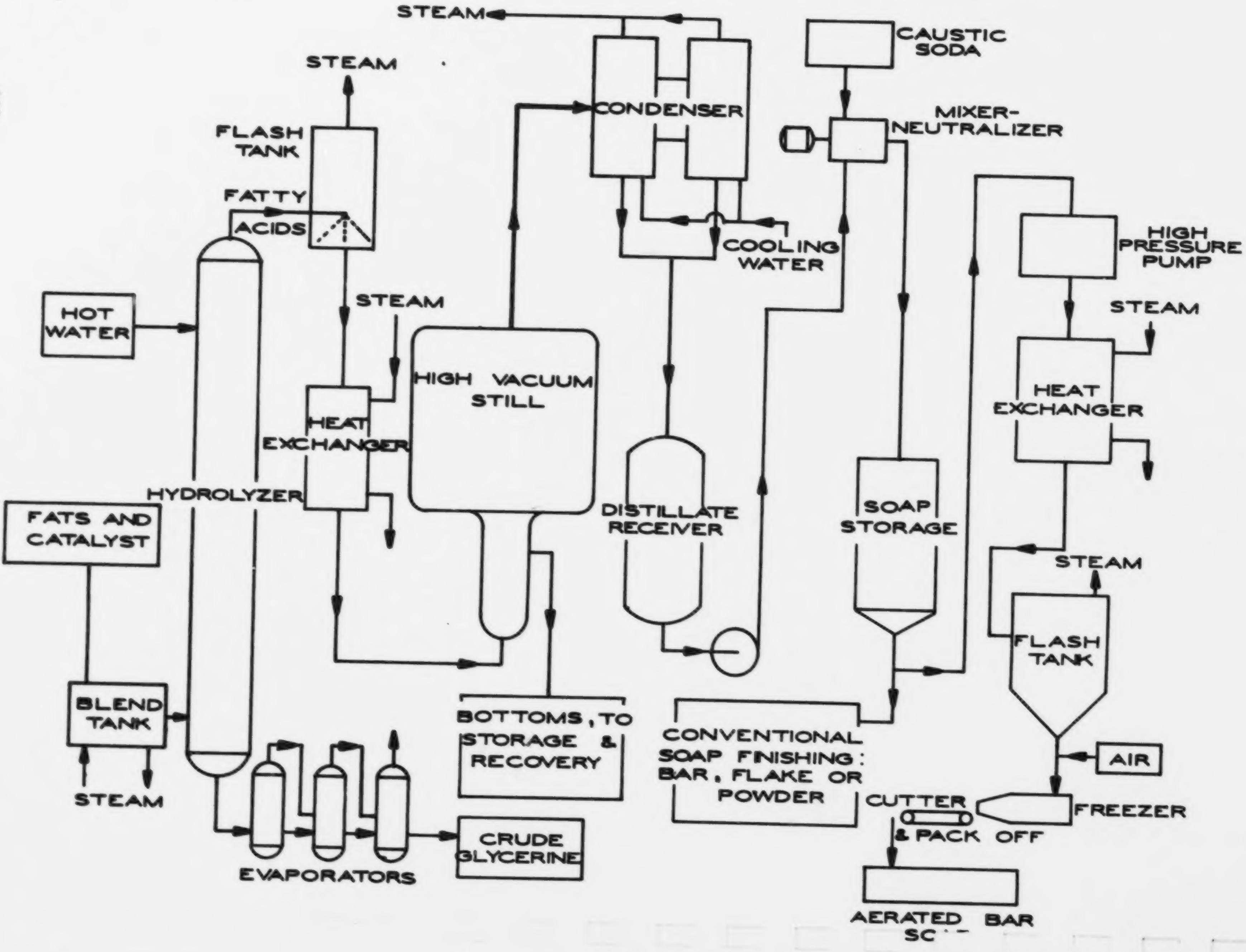
The manufacture of soap is a bit simpler. Soap consists of the sodium or potassium salts of various fatty acids, usually in the C_{12} to C_{18} range. Soap is usually manufactured either by saponification of fats or the hydrolysis of fats followed by neutralization of the resultant fatty acids. The trend today is toward the latter process, although direct saponification is much older and still widely used.

The main raw materials used in soap manufacture are tallow, greases, and coconut oil. These are the fatty materials which are hydrolyzed, or split, into fatty acids and glycerine. In some cases, however, free fatty acids are used as raw material and are neutralized directly. Fatty acids are usually produced from fats in one of three processes: Twitchell, batch autoclave, or continuous high-pressure process. The Twitchell process is the oldest. The fats are boiled for 12 to 48 hours at 212 to 220°F in an open tank with 0.75 to 1.25% Twitchell's reagent (alkyl-aryl sulfonic acids) and 0.5% sulfonic acid. Hydrolysis is 85 to 98% complete.(10, 11) A 5 to 15% aqueous glycerine solution is drawn off, neutralized and concentrated by evaporation.(10) The batch autoclave process is a faster process (5 to 10 hr) and operates at pressures from 75 to 150 psig and temperatures of 300 to 350°F in the presence of 1 to 2% zinc, calcium, or magnesium catalyst.(11) In the continuous countercurrent process, the fats (or fatty oil) are first deaerated under a vacuum. Deaerated fatty oil is charged to the bottom of the hydrolyzing tower. Deaerated water enters at the top of the tower. The oil at the bottom extracts the fatty material dissolved in the glycerine phase, while the water extracts the glycerine dissolved in the fatty phase. Final splitting occurs at about 485°F and 600 to 700 psig; direct high pressure steam is used for heating. The fatty acids exit through the top and the glycerine at the bottom.

The fatty acids produced in one of these processes (mainly the continuous countercurrent process) are now neutralized, usually with sodium hydroxide. This process is shown in Figure 125.(10) The crude fatty acids from the hydrolyzing tower first have the water flushed off, and then are purified by dry vacuum distillation. The distillate is then condensed and neutralized in a high speed mixer with a NaOH solution. The neat soap, or soap of the higher fatty acids, is discharged at temperature greater than 200°F to a blending tank. This soap contains 0.02 to 0.10 wt.% free caustic as Na_2O, 0.3 to 0.6% salt, and about 30% moisture. The soap is now dried in hot-air dryers, spray dryers, or steam heated tubes. Most soap is dried in hot-air dryers. Soap used in chips, bars and flakes is dried this way and then milled and broken into chips. The chips can be charged to mixers where other ingredients (perfume) are added. The chips are now finally rolled, and finished as bars or flakes.(11) Soap powders are produced by spray drying, which has been previously discussed.

FIGURE 125

CONTINUOUS PROCESS FOR SOAP MANUFACTURE


STEAM
STEAM
FLASH TANK
FATTY ACIDS
HOT WATER
STEAM
HEAT EXCHANGER
HYDROLYZER
FATS AND CATALYST
BLEND TANK
STEAM
EVAPORATORS
HIGH VACUUM STILL
BOTTOMS, TO STORAGE & RECOVERY
CRUDE GLYCERINE
CONDENSER
COOLING WATER
DISTILLATE RECEIVER
CAUSTIC SODA
MIXER-NEUTRALIZER
SOAP STORAGE
CONVENTIONAL SOAP FINISHING: BAR, FLAKE OR POWDER
HIGH PRESSURE PUMP
STEAM
HEAT EXCHANGER
STEAM
FLASH TANK
AIR
FREEZER
CUTTER & PACK OFF
AERATED BAR
SC

This entire process takes about 24 hours.

In contrast, the kettle or batch processes take several days to produce an equivalent amount of soap. The major batch process is the "full boiled process". This process employs direct saponification of fats, and consists of several steps, or "changes." The fats are first melted and mixed with 12.6% NaOH in the soap kettle, which is heated with open steam. After 3 to 4 hours, a sodium chloride solution is added until a concentration of 10 to 12% is achieved and the soap mixture separates distinctly into soap and spent lye. The steam is removed and the lye, containing 1 to 8% glycerine, is settled and is drawn off. The soap is again boiled and water is added until the mixture is again creamy; more caustic may be added and the process repeated. These series of processes are called the "brine changes." On the third day, the last of the oil and fats are saponified with 18.6% NaOH; the mixture is heated with closed steam until separation and then allowed to cool for 3 to 4 hours. The lye is run off and used in saponification of a new batch of fats; this step is called the "strengthening change." In the "finishing change," the soap is again heated and water is added, and the mixture is boiled. Upon cooling, the soap separates into 3 layers. The top layer is neat soap, the middle layer consists of nigre (a dark, alkaline soap), and the bottom layer is a mixture of soap and lye. The neat soap is finished as described above. The nigre is used either in the next batch or degraded and used in darker, cheaper grades of soap, and the lye is neutralized and recycled after the glycerine has been removed. The Sharples process is a continuous full-boiled process which takes 2 hours to produce neat soap. This process contains the same stages as the batch process, but the flow of lye is countercurrent to the flow of fat and oil. The soap, lye, and nigre are separated in 15,000 rpm centrifuges which develop a centrifugal force 13,200 times that of gravity.(10)

PROCESS CHEMISTRY

Many chemical reactions are involved in soap and detergent manufacture. The reactions involved in soap manufacture can be summarized as follows:(10)

1. Saponification of fats

$3\,NaOH + (C_{17}H_{35}COO)_3C_3H_5 \rightarrow$

Caustic Soda Glyceryl Stearate

$3\,C_{17}H_{35}COONa + C_3H_5(OH)_3$

Sodium Stearate Glycerine

2a. Hydrolysis of fats

$$(C_{17}H_{35}COO)_3C_3H_5 \;+\; 3H_2O \;\rightarrow\; 3C_{17}H_{35}COOH \;+\; C_3H_5(OH)_3$$

Glyceryl Stearate		Stearic Acid	Glycerine

b. Neutralization of fatty acids

$$C_{17}H_{35}COOH \;+\; NaOH \;\rightarrow\; C_{17}H_{35}COONa \;+\; H_2O$$

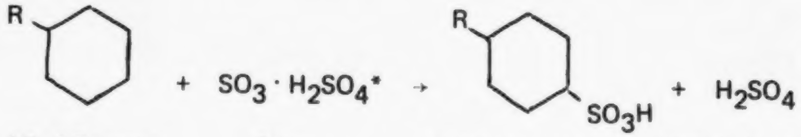
Stearic Acid	Caustic	Sodium Stearate

Although stearate has been shown in these reactions, other fats with carbon chains of 12 to 18 atoms, such as oleate, laurate, or palmitate are also present.

The formation of sulfonated detergents takes place mainly by the reactions summarized below.

1. Alkyl-benzene sulfonation:(2)

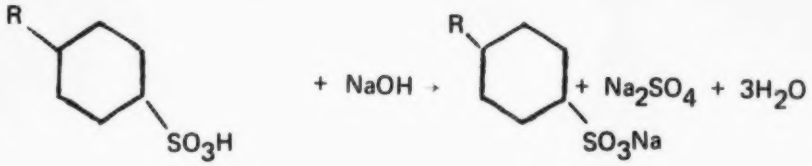
a. Sulfonation stage, 98+% conversion



Alkyl Benzene, C_{10} to C_{14} — Oleum — Sulfonic Acid — 96% Sulfuric acid

*An SO_3-air mixture is often used instead of oleum.

b. Neutralization



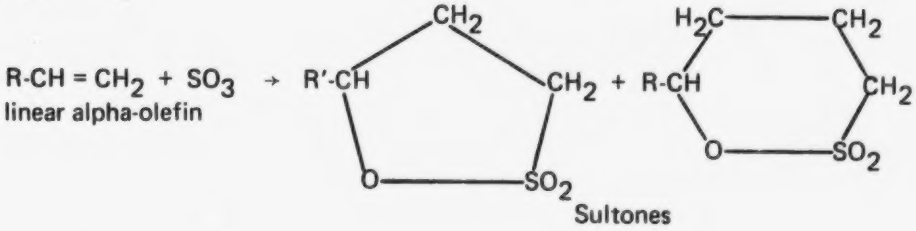
Sodium alkyl benzene sulfonate

2. Alpha-olefin sulfonation:(2)

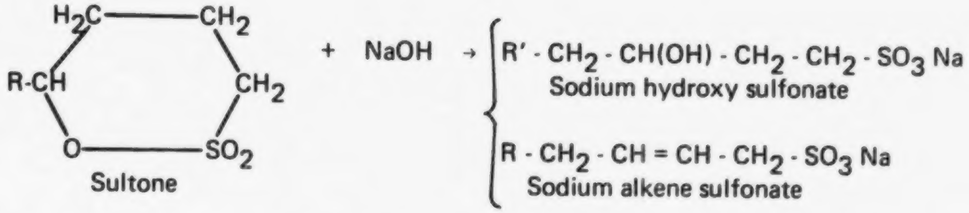
a. Sulfonation, 96 to 98+% conversion

$R\text{-}CH = CH_2 + SO_3 \rightarrow R\text{-}CH = CH\text{-}(CH_2)_n \text{-} SO_3H$

linear alpha-olefin, C_{14} to C_{18} — alkene sulfonic acid


R-CH = CH2 + SO3
linear alpha-olefin
R'-CH
CH2
CH2
O
SO2
H2C
CH2
R-CH
CH2
O
SO2
Sultones

b. Neutralization


H2C
CH2
R-CH
CH2
O
SO2
Sultone
+ NaOH
R' - CH2 - CH(OH) - CH2 - CH2 - SO3 Na
Sodium hydroxy sulfonate
R - CH2 - CH = CH - CH2 - SO3 Na
Sodium alkene sulfonate

$R' \text{-} CH = CH \text{-} (CH_2)_n \text{-} SO_3H + NaOH \rightarrow R' \text{-} CH = CH \text{-} (CH_2)_n \text{-} SO_3Na + H_2O$

Alkene sulfonic acid — Sodium alkene sulfonate

Composition of final product:(5)

50 to 60% alkene sulfonate
40 to 50% hydroxy sulfonate

3. Alcohol sulfation:

a. Sulfation(8)

I. With chlorosulfonic acid; 95 to 98% conversion(3)

$R\text{-}OH$	+	$ClSO_3H$	→	$R\text{-}O\text{-}SO_3H$	+	$HCl \uparrow$
Primary linear alcohol, C_{12} to C_{18}		Chlorosulfonic acid		Alkyl sulfuric acid		

II. With sulfur trioxide, 95 to 98% conversion

$$\underset{\text{excess}}{R\text{-}OH + \underset{3.5\%}{SO_3}} \rightarrow R\text{-}O\text{-}SO_3H$$

III. With sulfuric acid (oleum or monohydrated), 65% conversion

$$R\text{-}OH + \underset{\substack{70\% \\ \text{excess}}}{H_2SO_4} \rightleftarrows R\text{-}O\text{-}SO_3H + H_2O$$

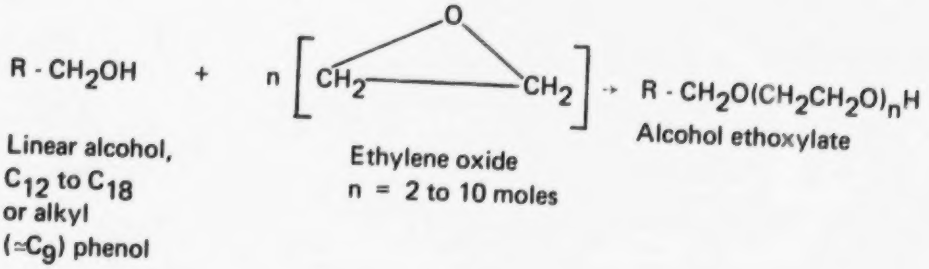
b. Neutralization

I. $R\text{-}O\text{-}SO_3H + NaOH \rightarrow \underset{\text{Sodium alcohol sulfate}}{R\text{-}O\text{-}SO_3Na} + H_2O$

II. $R\text{-}O\text{-}SO_3H + H_2SO_4 + 3NaOH \rightarrow \underset{55\%}{R\text{-}O\text{-}SO_3Na} + \underset{45\%}{Na_2SO_4} + 3H_2O$

4. Alcohol ethoxylation and sulfation:(6, 10)

a. Ethoxylation


R - CH2OH
+
n
O
CH2
CH2
R - CH2O(CH2CH2O)nH
Alcohol ethoxylate
Linear alcohol,
C12 to C18
or alkyl
(≈C9) phenol
Ethylene oxide
n = 2 to 10 moles

b. Sulfation and Neutralization

$$\underset{\substack{\text{Alcohol ethoxylate} \\ R = C_{11} \text{ to } C_{17} \\ n = 2 \text{ to } 4}}{R\text{-}CH_2O(CH_2CH_2O)_nH} + SO_3 \rightarrow R\text{-}CH_2O(CH_2CH_2O)_nSO_3H$$

$$R\text{-}CH_2O(CH_2CH_2O)_nSO_3H + NaOH \rightarrow \underset{\text{Alcohol ethoxy sulfate}}{R\text{-}CH_2O(CH_2CH_2O)_nSO_3Na} + H_2O$$

RAW MATERIALS AND PRODUCTS

The raw materials for soap manufacturing consist mainly of tallow, greases, coconut oil, and free fatty acids. Tallow is obtained mainly by the steam rendering of cattle fat; greases are obtained from the steam rendering of the fats of hogs, smaller domestic animals, and from garbage.

An extremely wide range of compounds, both natural and synthetic may be classified as raw materials for the manufacture of synthetic detergents. The active ingredient in detergents comprises about 14 to 37% of the total product; the remainder is composed of builders, fillers, etc., which are not generally produced by the detergent manufacturers. Table 159 shows typical formulations of modern detergent products.(3)

The basic raw material used in the manufacture of LAS is a linear alkyl benzene, mainly dodecyl alkyl benzene. This alkylated benzene is produced by either dehydrogenation or chlorination of a n-paraffin, followed by Friedel-Craft alkylation.

Alpha-olefin sulfonate is made by employing a linear paraffin wax or ethylene to produce a C_{15} to C_{18} alpha-olefin, which is then sulfonated. In the production of alcohol sulfates or ethoxylated sulfates, straight chain primary alcohols (C_{12} to C_{18}) are usually the feed material. Lauryl alcohol, derived from coconut oil, is often used, but it is being replaced by synthetic alcohols made by either ethylene growth over Ziegler catalysts or the reaction of olefins with carbon monoxide and then hydrogen.(6)

The major materials used in sulfonation or sulfation are, as previously mentioned, stabilized liquid sulfur trioxide, oleum, chlorosulfonic acid, or sulfur. Alcohol ethoxylates are prepared by reacting ethylene oxide or propylene oxide with linear alcohols. Neutralization of the sulfonic acid is carried out by an agent which does not produce gaseous side products. Sodium hydroxide is generally used, but potassium, lithium, magnesium, and ammonium hydroxides, as well as trietanolamine are also employed.

Soaps and synthetic detergents are available in powder, liquid, bar, flake and chip form. They are used primarily as cleaning compounds, and have a large range of home and industrial applications. Major products include laundry detergents, toilet soap, and dish washing detergents used in the home. Detergents are also used in textile finishing, dyeing, laundries, dairies, and metal cleaning. They are used as a frothing agent in wallboard manufacture and as an entraining agent in portland cement. Soap is used in the manufacture of

TABLE 159

TYPICAL SYNTHETIC DETERGENT FORMULAS[3]

	Heavy-duty product U.S., Canada	Light-duty powders all countries	Light-duty liquids all countries	Detergent laundry bars
Active ingredient (e.g., alkylbenzene sulfonate, fatty alcohol sulfate)	14 to 20%	25 to 32%	30 to 37%	20 to 25%
Foam booster (e.g., lauryl alcohol, cocomonoethanolamide)	1.5 to 2%		5 to 12%	
Sodium tripolyphosphate	40 to 60%	2 to 15%		15 to 25%
Anti-soil redeposition agent (e.g., sodium carboxy methyl cellulose)	0.5 to 0.9%			0.3 to 0.5%
Anti-corrosion agent (e.g., sodium silicate)	5 to 7%	0.02 to 0.08%		3 to 8%
Optical brightener	0.30 to 0.75%			0.05 to 0.25%
Enzymes	0.20 to 0.75%			
Moisture	6 to 12%	1 to 4%		3 to 8%
Filler (e.g., sodium sulfate)	remainder	60 to 68%		15 to 30%
Other ingredients			6 to 9%	15 to 20%

synthetic rubber, textile processing and as an emulsifier in cold cream and cosmetic preparations. Glycerine, which is a major by-product of soap manufacture, is used in the manufacture of alkyd resins and explosives, as a humectant in tobacco, as a plasticizer in cellophane, and in products such as cosmetics, dentrifices, corks, gaskets, and pharmaceuticals.(10)

NATURE OF THE GASEOUS DISCHARGE

In soap manufacturing plants, particulate and/or odors are the main atmospheric contaminant. The odors are mostly due to the storage and unloading of the fat charge stock particularly rendered fats and greases, and the vacuum distillation of higher molecular weight fatty acids. Excepting spray dryers, dust emissions in soap making cause little more than dust problems within the plant.

The main source of air pollution in a detergent plant is the exhaust air from the spray dryer. Dust from dry product handling and acid fumes from the sulfonation process can also be problems.

The exhaust air from a countercurrent spray dryer will be at a temperature of 150 to 250°F, typically about 210 to 220°F. The air is exhausted to a collector, either a cyclone or a wet scrubber, and the collected particles (in the form of fines or in a slurry) are recycled to the process. The spray dryer is essentially an evaporator, reducing the moisture content from about 40% to between 6 and 15%. The wet bulb temperature of the exhaust gas before entering the collector will be 120 to 150°F and the particulate dust loading will be in the 3 to 7 gr/SCF range. This particulate content will depend on the tower rates, the fineness of the sprays, the characteristics of the specific product being dried and other variables. The particles will be rather large, approximately 50% being greater than 40 microns.(13) The composition of the particles will be very close to that of the finished product, with the exception that sodium perborate, enzymes, and organic foam boosters are not present. A problem is caused by the great amount of moisture evaporated, which is discharged in the exhaust gas and condenses in the atmosphere, forming a dense white plume.

Detergent formulations have recently been undergoing changes in response to public concern over the possible connection of phosphates to algae growth in inland waterways. As a result, the industry trend has been toward non-phosphate formulations. It is speculation as to the make-up of the new detergents, but the emphasis appears to be on non-ionic surfactants such as alcohol ethoxylates.(14) At any rate, these new formulations produce problems in spray dryer emission control. The spray dryer emits a blue haze which has not been controlled by conventional technology. This haze is believed to be a sub-micron organic aerosol mist. As the scope of this contract is limited to conventional emission control technology, only the phosphate detergent pollution problem will be dealt with in detail.

The pneumatic air lift used to cool and convey the dried product contains the potential for sizable particulate emissions. Particulate rates from the separating device at the top of the lift average about 3 to 4% of the amount of beads conveyed. The air used in the pneumatic lift will be approximately 1 cfm/lb/hr of product and will have a temperature of near 100° F.(13)

There are various gaseous contaminants released during the sulfonation processes. In the burning of sulfur to produce the SO_3-air sulfonating mixture, some of the SO_2 is not converted to SO_3. This will result in a small amount of sulfur dioxide emission. If oleum is used as the sulfonating agent, a small amount of acid vapor will escape from the unloading, storage, sulfonating, aging, and neutralization equipment. When chlorosulfonic acid is used in the sulfation process, large quantities of gaseous hydrogen chloride are produced and must be disposed of.

POLLUTION CONTROL CONSIDERATIONS

The control of odors in a soap and detergent manufacturing plant, compared to the control of product particulate, is a rather simple problem. The majority of odor emissions in soap making plants will come from the vacuum type steam stills in the neutralization process, or the soap kettles in the full-boiled process. These odors may be controlled by spray condensers or surface condensers and the noncondensibles vented to the firebox of a continuous boiler.(13) The odor level from detergent spray dryers will be about four or five odor units/SCF. Although usually this will not constitute a problem, discharging the exhaust gas at a high velocity from a tall stack will greatly lessen the chance of the plume reaching ground in any noticeable concentration.(13)

In the control of particulate emissions from spray dryers, cyclones are frequently employed, as the fines from the dryer amount to an economically recoverable portion. The cyclones used are high efficiency (90% or better) and relatively high pressure drop (4 to 5 in. w.c.), and are used alone or with one or more additional units in series. Two-stage cyclone collectors operating at 12 in.

w.c. are reported to obtain collection efficiencies of 99%. The small tube type of cyclone is not usually employed. The cyclones are usually carefully insulated to prevent solids build-up.

The gases discharged from a single cyclone will generally have a particulate loading of 0.3 to 0.6 gr/SCF. Much of this weight will be accounted for by a few large particles which escape the cyclone.(13) This level of particulate loading is not economically recoverable, but may be in violation of many emission standards. Therefore, some form of subsequent control equipment will be required. A two-stage cyclone will have emission loadings in the order of 0.1 gr/SCF.

Problems may be encountered when fabric filters are used as control devices for detergent drying towers, as the exhaust gas is very humid and the product particles are sticky at high temperatures. These conditions can cause caking and build-up problems in the filter.

Wet scrubbing seems to offer the best means of reducing the particulate loading of the cyclone exhaust gas. Spray chambers, centrifugal impingement scrubbers, and Venturi scrubbers have all been used. Velocities in spray chambers should be a maximum of 250 ft/min to prevent water entrainment; scrubbing liquid rates as low as 1 to 2 gpm for every 1000 cfm of exhaust are satisfactory. Emission concentrations of about 0.15 gr/SCF are normal for spray chambers. Venturi scrubbers would use a scrubbing liquid rate of about 8 to 10 gpm/1000 acfm. A medium pressure drop (8 to 10 in. w.c.) Venturi scrubber would discharge air with particulate loadings as low as 0.02 to 0.03 gr/SCF. If a low gas pressure drop (3 to 5 in. w.c.) were used, the particulate loading would be about 0.1 gr/SCF.(13) To prevent foaming in the scrubber, a high detergent concentration should be maintained in the recirculated liquid; 40 to 45 wt.% solids is a typical range.

The emissions from the pneumatic air lift represent an economically important fraction of the product. This 3 to 4% fraction is equivalent to a grain loading of 3.5 to 4.7 gr/ACF in the separator exhaust. The usual method of recovering these fines is the use of a fabric filter. Filter caking is not as much of a problem here as the detergent beads are less sticky at lower temperatures.

SPECIFICATIONS AND COSTS

Abatement system specifications have been written to control the emissions from two sources in a detergent manufacturing plant. The first of these sources is the spray dryer vent. Both wet scrubber and fabric filter systems have been specified. Each system is based upon the presumption that the detergent being made is of the conventional high phosphate variety and, therefore, no hydrocarbon mist problem exists. Cost data and equipment specifications for these two systems are presented in Tables 160 to 170 and Figures 126 to 134.

The second source for which specifications have been written is the dry product handling operation which transfers spray dried product from the spray dryer to the storage facility. Transfer is done by pneumatic conveying in which the air both transports and cools the detergent product. The single abatement device specified for this application is the fabric filter. Cost data and specifications are presented in Tables 171 to 175 and Figures 135 to 137.

Preceding page blank

TABLE 160

WET SCRUBBER PROCESS DESCRIPTION FOR SOAP AND DETERGENT SPRAY DRYING SPECIFICATION

The wet scrubber is to remove entrained detergent particles from the exhaust gas of a gas-fired spray dryer producing conventional phosphate detergents. The flow of air and detergent in the spray dryer is countercurrent. Detergent product is collected at the bottom of the spray drying tower and sent to storage by a conveyor system. The exhaust gases leave the tower at the top.

The exhaust gas from the spray dryer will be brought to a fan located outside of the building. The fan outlet is to be five feet above grade. The wet scrubber will be located beyond the fan in an area free of space limitations. The scrubber is to be designed to withstand the discharge pressure developed by the fan. The ducting to the scrubber should be insulated to alleviate build-up.

The scrubber will recirculate a slurry of 40 to 45 wt.% solids content in order to minimize foaming problems in the unit. Fresh water is available at the site for make-up. Slurry withdrawn from the unit will be returned to the detergent manufacturing process. The temperature of the slurry in the recycle tank should be kept in the range of approximately 130°F.

The scrubbing system will include:

1. *A wet scrubber equipped with suitable connections for periodic cleaning of solids build-up.*
2. *Carbon steel construction with manufacturer's standard prime paint applied.*
3. *A fan capable of overcoming the system pressure drop at the design flow rate while operating at no more than 90% of the maximum recommended speed. The fan motor shall be capable of driving the fan at the maximum recommended speed and the corresponding pressure differential at 20% over the design flow rate.*
4. *Pumps capable of handling slurrys at the design solids content.*
5. *A fifty foot stack following the scrubber.*

All equipment other than the scrubber should be considered auxiliary equipment.

TABLE 161

WET SCRUBBER OPERATING CONDITIONS FOR SOAP AND DETERGENT SPRAY DRYING SPECIFICATIONS

Two sizes of wet scrubbers are to be quoted for each of two efficiency levels. Vendors' quotations should consist of four separate and independent quotations.

	Small	*Large*
Dryer Capacity, lb/hr	*20,000*	*80,000*
Process Wt., lb/hr	*30,000*	*120,000*
Inlet Gas to Scrubber		
Flow, ACFM	*30,290*	*121,150*
Temperature, °F	*210*	*210*
Flow, SCFM	*23,960*	*95,840*
% Moisture (vol)	*15*	*15*
Solids Rate, lb/hr	*1,298*	*5,192*
Solids Loading, gr/DSCF	*7.44*	*7.44*
Scrubber Outlet Gas		
Volume, ACFM	*27,450*	*109,800*
Temperature, °F	*134*	*134*
% Moisture (vol)	*17*	*17*
	Case 1 – Medium Efficiency*	
Outlet Solids Rate, lb/hr	*25.2*	*40.0*
Outlet Solids Loading, gr/DSCF	*0.14*	*0.06*
Efficiency, wt.%	*98.04*	*99.23*
Scrubber, ΔP	*10 in. w.c.*	*10 in. w.c.*
	Case 2 – High Efficiency*	
Outlet Solids Rate, lb/hr	*2.35*	*9.41*
Outlet Solids Loading, gr/ACF	*0.01*	*0.01*
Outlet Solids Loading, gr/DSCF	*0.0135*	*0.0135*
Collection Efficiency, wt.%	*99.82*	*99.82*
Scrubber, ΔP	*40 in. w.c.*	*40 in. w.c.*

**See page 6 for definition of efficiency levels.*

TABLE 162

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING

	Medium Efficiency		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	30,290	121,150	30,290	121,150
°F	210	210	210	210
SCFM	23,960	95,840	23,960	95,840
Moisture Content, Vol. %	15	15	15	15
Effluent Contaminant Loading				
gr/ACF	5.00	5.00	5.00	5.00
lb/hr	1,298	5,192	1,298	5,192
Cleaned Gas Flow				
ACFM	27,450	109,800	27,450	109,800
°F	134	134	134	134
SCFM	24,492	97,970	24,492	97,970
Moisture Content, Vol. %	17	17	17	17
Cleaned Gas Contaminant Loading				
gr/ACF	0.11	0.04	0.01	0.01
lb/hr	25.2	40.0	2.35	9.41
Cleaning Efficiency, %	98.0	99.3	99.8	99.8
(1) Gas Cleaning Device Cost	28,500	69,300	28,000	62,650
(2) Auxiliaries Cost				
(a) Fan(s)	13,750	34,100	22,275	53,650
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering	52,300	94,400	67,050	124,450
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost	94,550	197,800	117,325	240,750

TABLE 163

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING

Operating Cost Item	Unit Cost	Medium Efficiency		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year		8,000	8,000	8,000	8,000
Operating Labor (if any)					
Operator					
Supervisor					
Total Operating Labor		1,800	1,800	1,800	1,800
Maintenance					
Labor					
Materials					
Total Maintenance		3,975	5,660	5,100	7,500
Replacement Parts					
Total Replacement Parts		1,825	2,925	2,485	4,005
Utilities					
Electric Power	$.011/kw-hr	5,199	38,500	22,880	86,790
Fuel					
Water (Process)	$.25/M gal	1,140	4,560	2,578	9,660
Water (Cooling)					
Chemicals, Specify					
Total Utilities		6,339	43,060	25,458	96,450
Total Direct Cost		13,939	53,445	34,843	109,755
Annualized Capital Charges		9,455	19,780	11,733	24,075
Total Annual Cost		23,394	73,225	46,576	133,830

FIGURE 126

CAPITAL COSTS FOR WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING (MEDIUM EFFICIENCY)

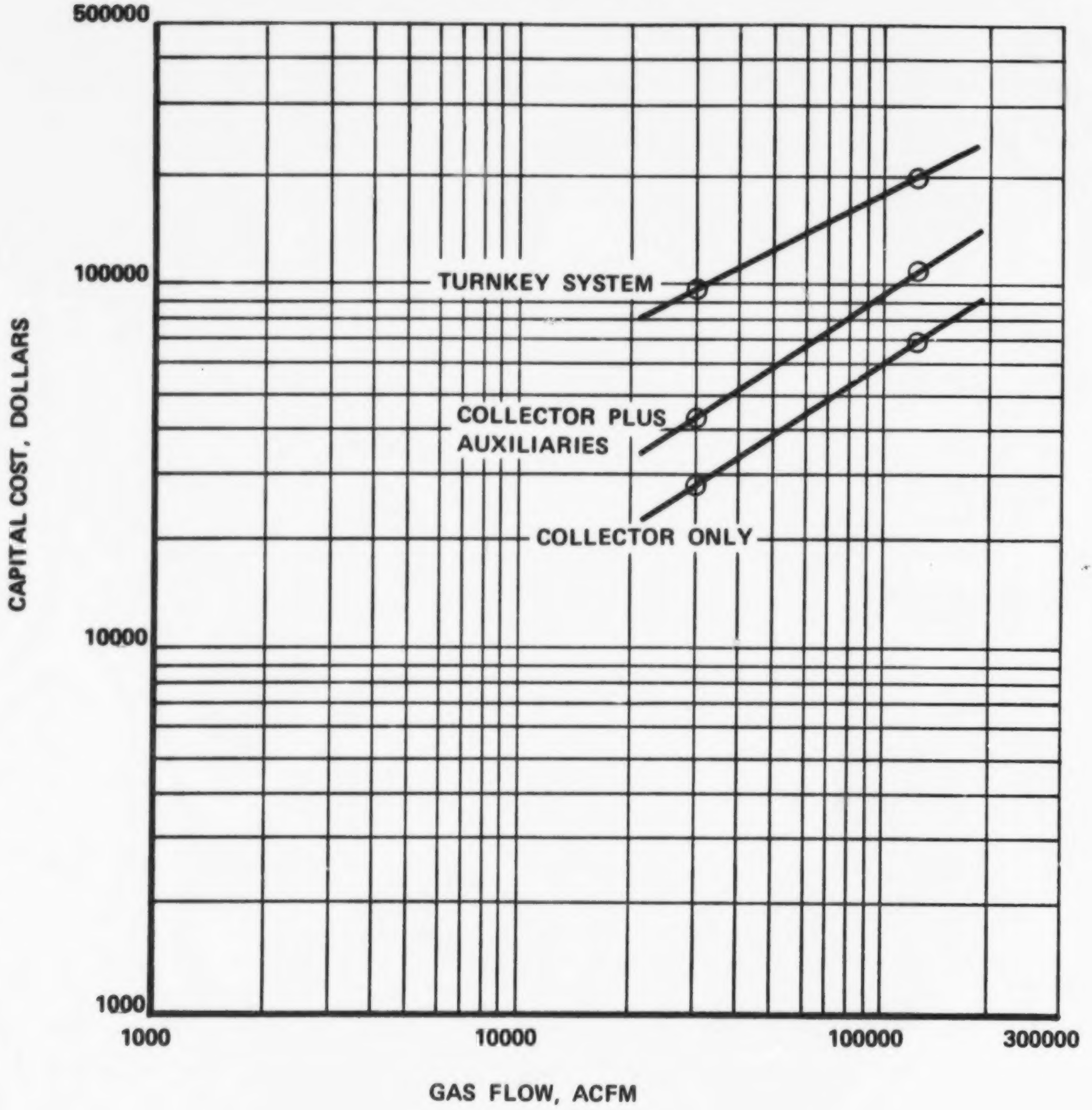


FIGURE 127

ANNUAL COSTS FOR WET SCRUBBERS
FOR SOAP AND DETERGENT SPRAY DRYING
(MEDIUM EFFICIENCY)

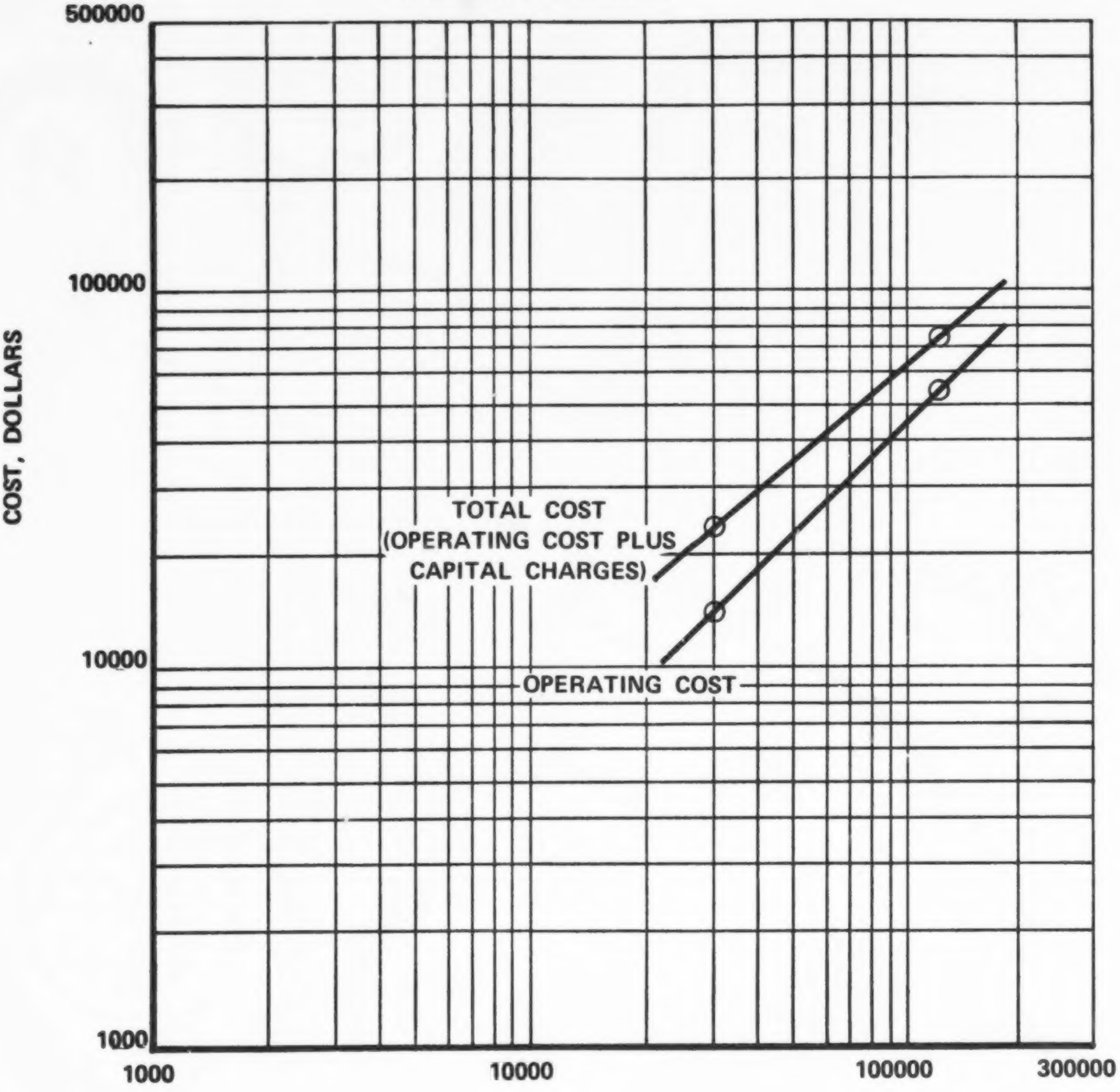


FIGURE 128

CAPITAL COSTS FOR WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING (HIGH EFFICIENCY)

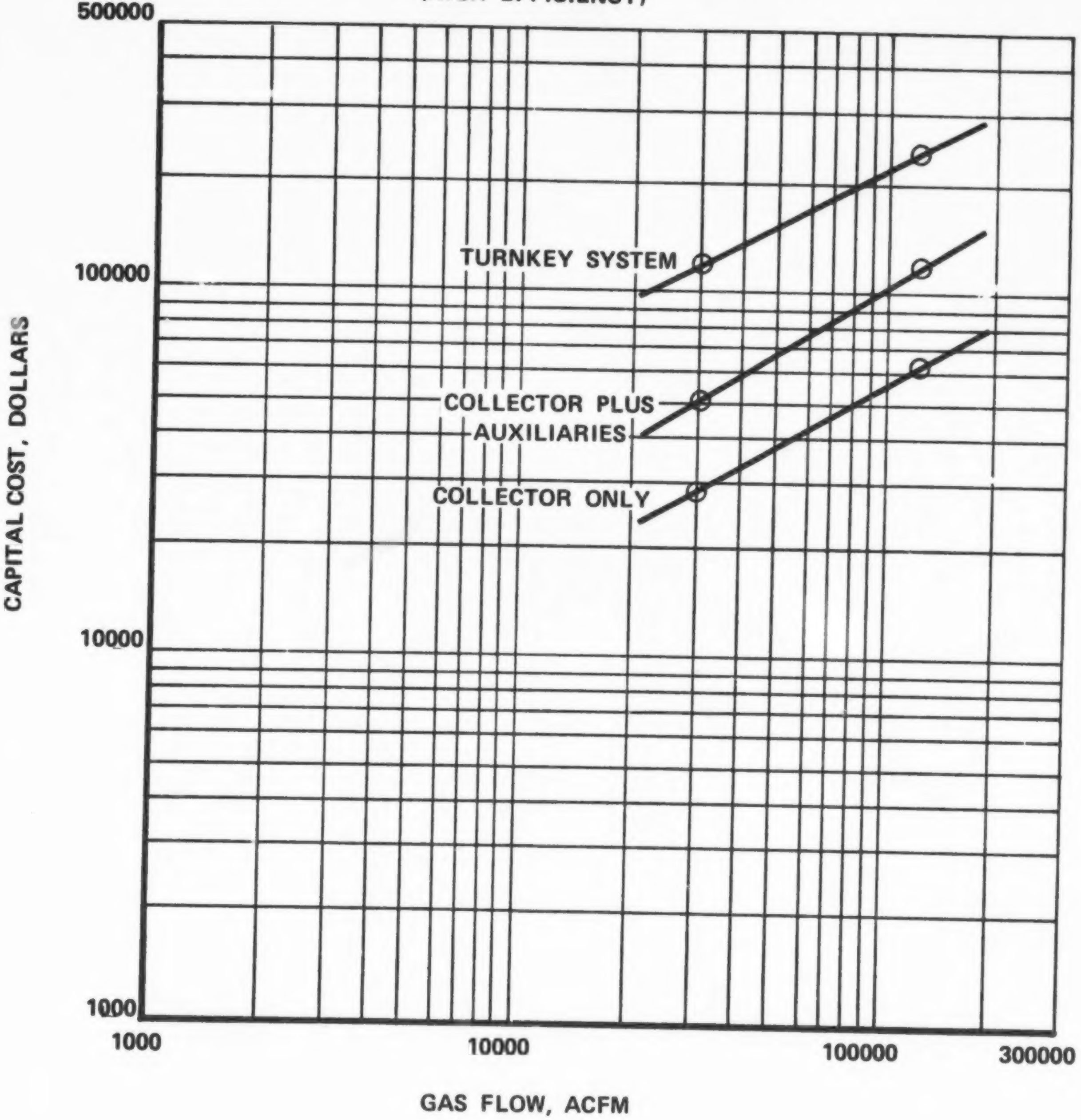


FIGURE 129

ANNUAL COSTS FOR WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING (HIGH EFFICIENCY)

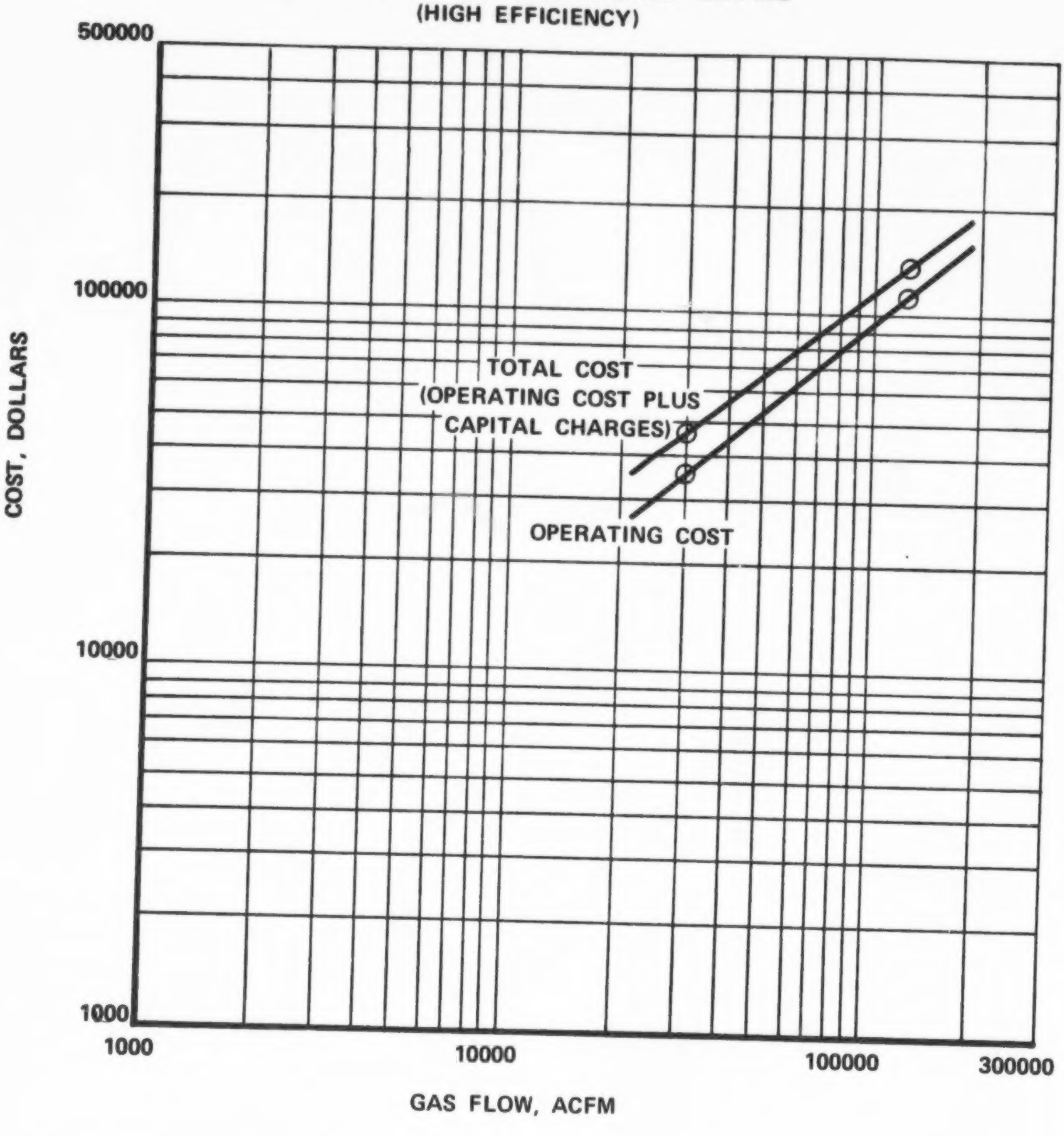


TABLE 164

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING (MEDIUM EFFICIENCY)

Population Size – 20 Sample Size – 2

Capital Cost = $94,550

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$ 90,142	$ 98,958
75	84,787	104,313
90	74,307	114,793
95	62,987	126,113

Capital Cost = $197,800

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$168,127	$227,473
75	132,086	263,514
90	61,540	334,060

FIGURE 130

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING (MEDIUM EFFICIENCY)

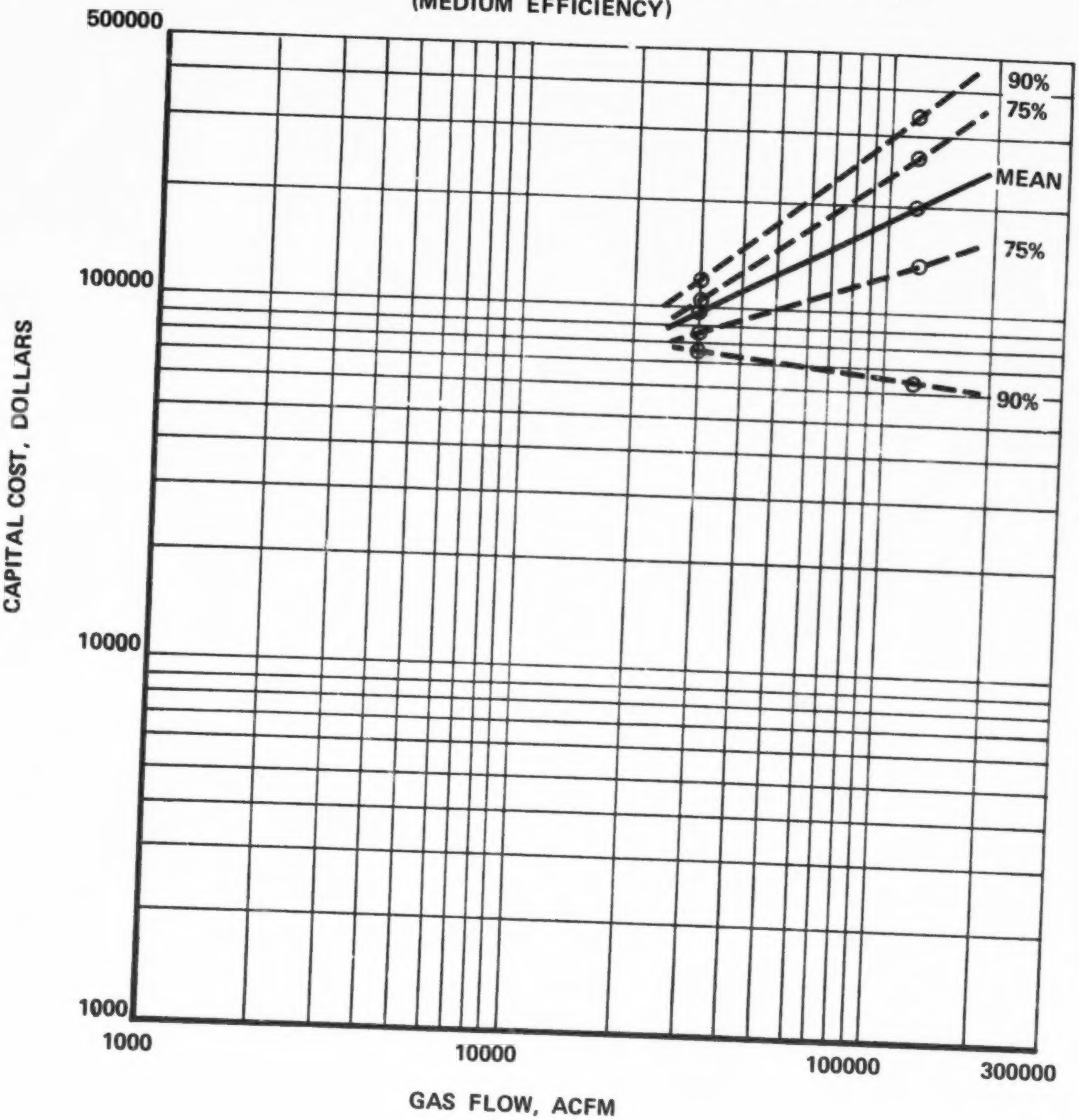


TABLE 165

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING (HIGH EFFICIENCY)

Population Size – 20 Sample Size – 2

Capital Cost = $117,325

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$92,036	$142,614
75	61,321	173,329
90	1,200	233,450

Capital Cost = $240,600

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$239,083	$242,117
75	237,241	243,959
90	233,635	247,565
95	229,740	251,460

FIGURE 131

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR SOAP AND DETERGENT SPRAY DRYING (HIGH EFFICIENCY)

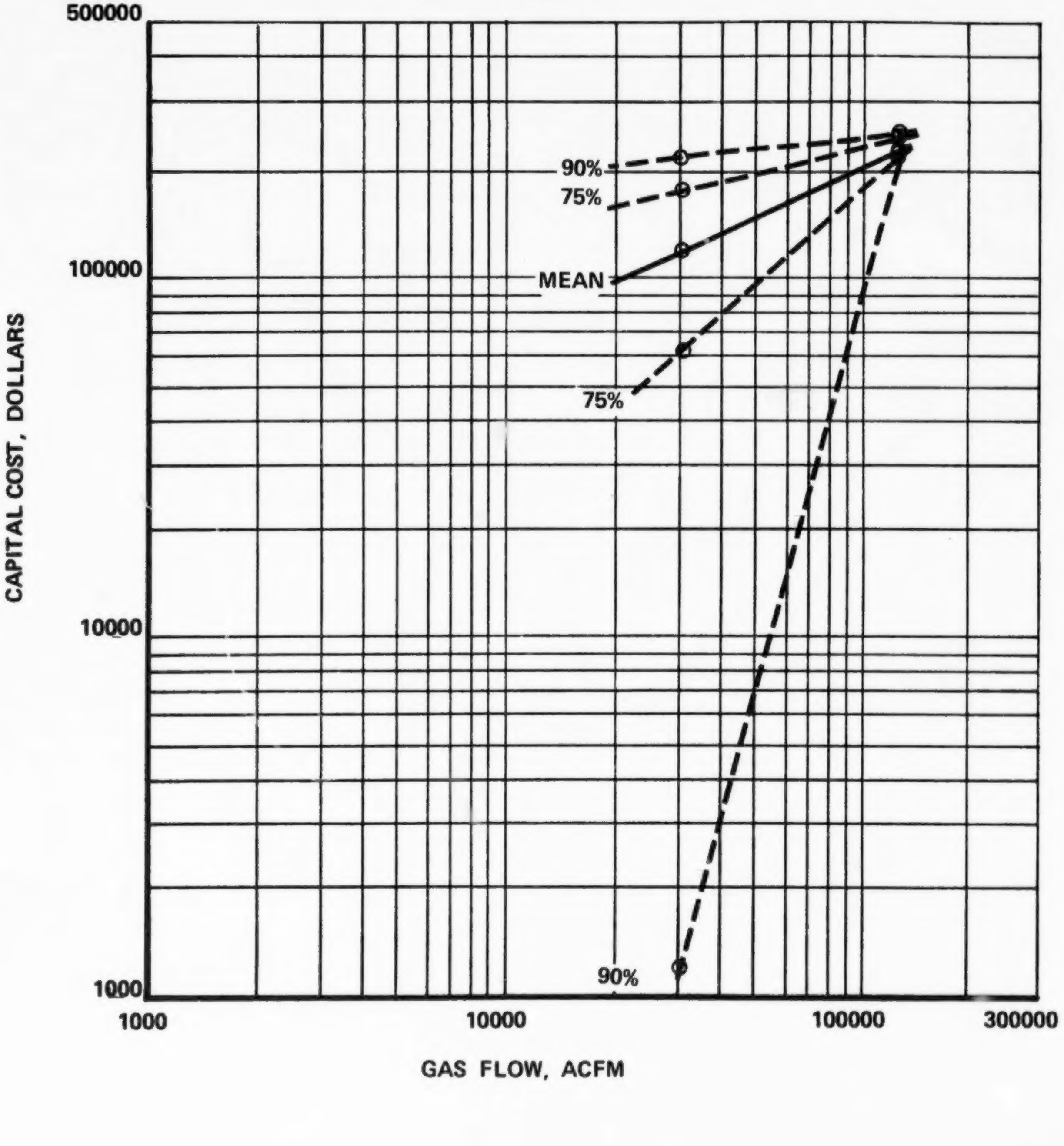


TABLE 166

FABRIC FILTER PROCESS DESCRIPTION FOR SOAP AND DETERGENT SPRAY DRYING SPECIFICATION

The fabric filter is to remove entrained detergent particles from the exhaust gas of a spray dryer processing conventional phosphate detergents. The flow of air and detergent in the dryer is countercurrent. The dryer is fired by natural gas. Before entering the filter, the dryer exhaust gas passes through one or more high efficiency cyclones with a combined efficiency of 90%. The recovered fines are returned to the process.

The exhaust gas from the cyclones will be brought to the filter by means of a fan. The fabric filter is to be located outside of the building in an area free of space limitations. The fan is to follow the filter and discharge into the base of a thirty foot stack. The ducting to the filter should be insulated to alleviate build-up.

The fabric filter is to consist of a minimum of five compartments and will allow for isolation of an individual compartment for cleaning during operation. Any four of these compartments should be capable of handling the full flow at designed differential pressure. Each section should also be capable of isolation for maintenance and have provisions for personal safety during filter operation. The collecting process is to be continuous and should include the following:

1. *Pulse jet fabric filter or equal, fully insulated, top removal design, compartmented as described above, with a maximum air-to-cloth ratio of 6/1 and a design differential pressure of 6 in. w.c.*

2. *Insulated hoppers with a minimum side angle of 60° and a manual slide gate at the screw conveyor for isolation. The hopper should be strong enough to support a full load although in routine operation the hopper will not be full.*

3. *Insulate and heat trace entire system ductwork.*

4. *Carbon steel construction with manufacturer's standard prime paint applied.*

5. *An insulated screw conveyor equipped with a single rotary air lock.*

6. *A fan, located downstream of the filter sized with 20% excess capacity when operating at the design pressure and 90% of the maximum recommended speed.*

TABLE 167

FABRIC FILTER OPERATING CONDITIONS FOR SOAP AND DETERGENT SPRAY DRYING SPECIFICATION

Two sizes of fabric filters are specified for the "high efficiency" level. Vendors' quotations should consist of one quotation for each of the two sizes, with a representation of the efficiency expected for the unit quoted. The efficiency quoted may be better than the "high efficiency" case.

	Small	*Large*
Dryer Capacity, lb/hr	*20,000*	*80,000*
Process wt, lb/hr	*30,000*	*120,000*
Inlet Gas to Primary Collector		
Flow, ACFM	*30,290*	*121,150*
Temperature, °F	*210*	*210*
Flow, SCFM	*23,960*	*95,840*
% Moisture (vol)	*15*	*15*
Inlet Solids Rate, lb/hr	*616*	*2,464*
Inlet Solids Loading, gr/ACF	*2.37*	*2.37*
Inlet Gas to Fabric Filter		
Flow, ACFM	*29,840*	*119,350*
Temperature, °F	*200*	*200*
Flow, SCFM	*23,960*	*95,840*
% Moisture (vol)	*15*	*15*
Dew Point, °F	*132*	*132*
Inlet Solids Rate, lb/hr	*61.6*	*246.4*
Inlet Solids Loading, gr/ACF	*0.24*	*0.24*
Inlet Solids Loading, gr/DSCF	*0.35*	*0.35*
	High Efficiency	
Outlet Solids Rate, lb/hr	*2.56*	*10.23*
Outlet Solids Loading, gr/ACF	*0.01*	*0.01*
Outlet Solids Loading, gr/DSCF	*0.015*	*0.15*
Collection Efficiency, wt.%	*95.94*	*95.85*
Air-To-Cloth Ratio	*6/1*	*6/1*

TABLE 168

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR FABRIC FILTERS FOR SOAP AND DETERGENT SPRAY DRYING

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			30,290	121,150
°F			210	210
SCFM			23,960	95,840
Moisture Content, Vol. %			15	15
Effluent Contaminant Loading				
gr/ACF			2.37	2.37
lb/hr			616	2,464
Cleaned Gas Flow				
ACFM			30,290	121,150
°F			210	210
SCFM			23,960	95,840
Moisture Content, Vol. %			15	15
Cleaned Gas Contaminant Loading				
gr/ACF			0.01	0.01
lb/hr			2.56	10.23
Cleaning Efficiency, %			95.94*	95.85*
(1) Gas Cleaning Device Cost			35,278	138,250
(2) Auxiliaries Cost				
(a) Fan(s)			11,558	31,293
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			4,905	13,993
(b) Foundations & Support			35,375	126,393
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost			87,116	309,929

*Cleaning efficiency for baghouse only; preceeded by 90% efficient cyclone.

FIGURE 132

CAPITAL COSTS FOR FABRIC FILTERS
FOR SOAP AND DETERGENT SPRAY DRYING

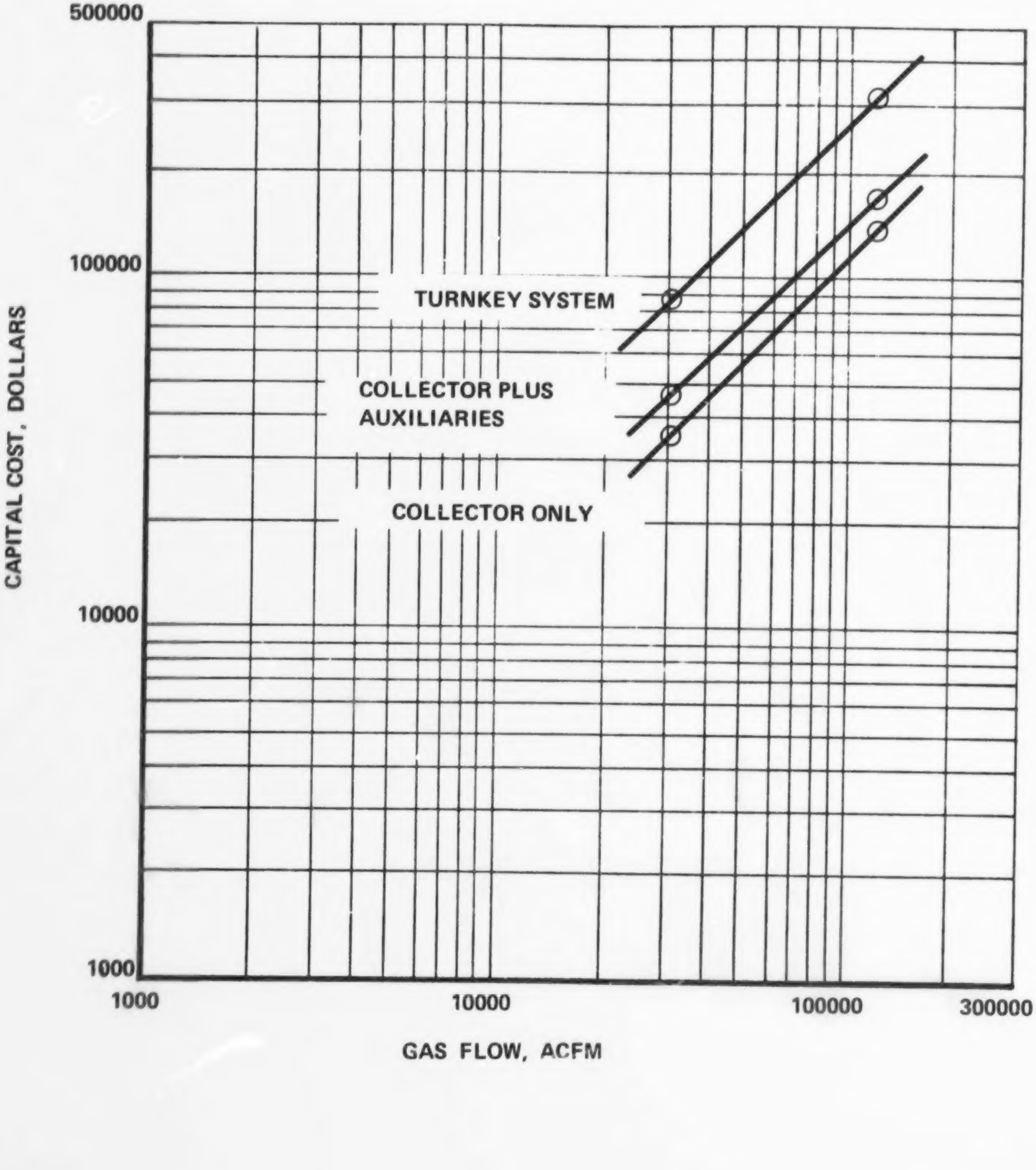


TABLE 169

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR FABRIC FILTERS FOR SOAP AND DETERGENT SPRAY DRYING

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				8,000	8,000
Operating Labor (if any)					
Operator					
Supervisor	$8/hr			290	290
Total Operating Labor				290	290
Maintenance					
Labor	$6/hr			1,300	4,298
Materials				719	2,597
Total Maintenance				2,019	6,895
Replacement Parts				1,075	3,600
Total Replacement Parts				1,075	3,600
Utilities					
Electric Power	$.011/kw-hr			3,932	17,508
Fuel					
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				3,932	17,508
Total Direct Cost				7,316	28,293
Annualized Capital Charges				8,712	30,993
Total Annual Cost				16,028	59,286

FIGURE 133

ANNUAL COSTS FOR FABRIC FILTERS FOR SOAP AND DETERGENT SPRAY DRYING

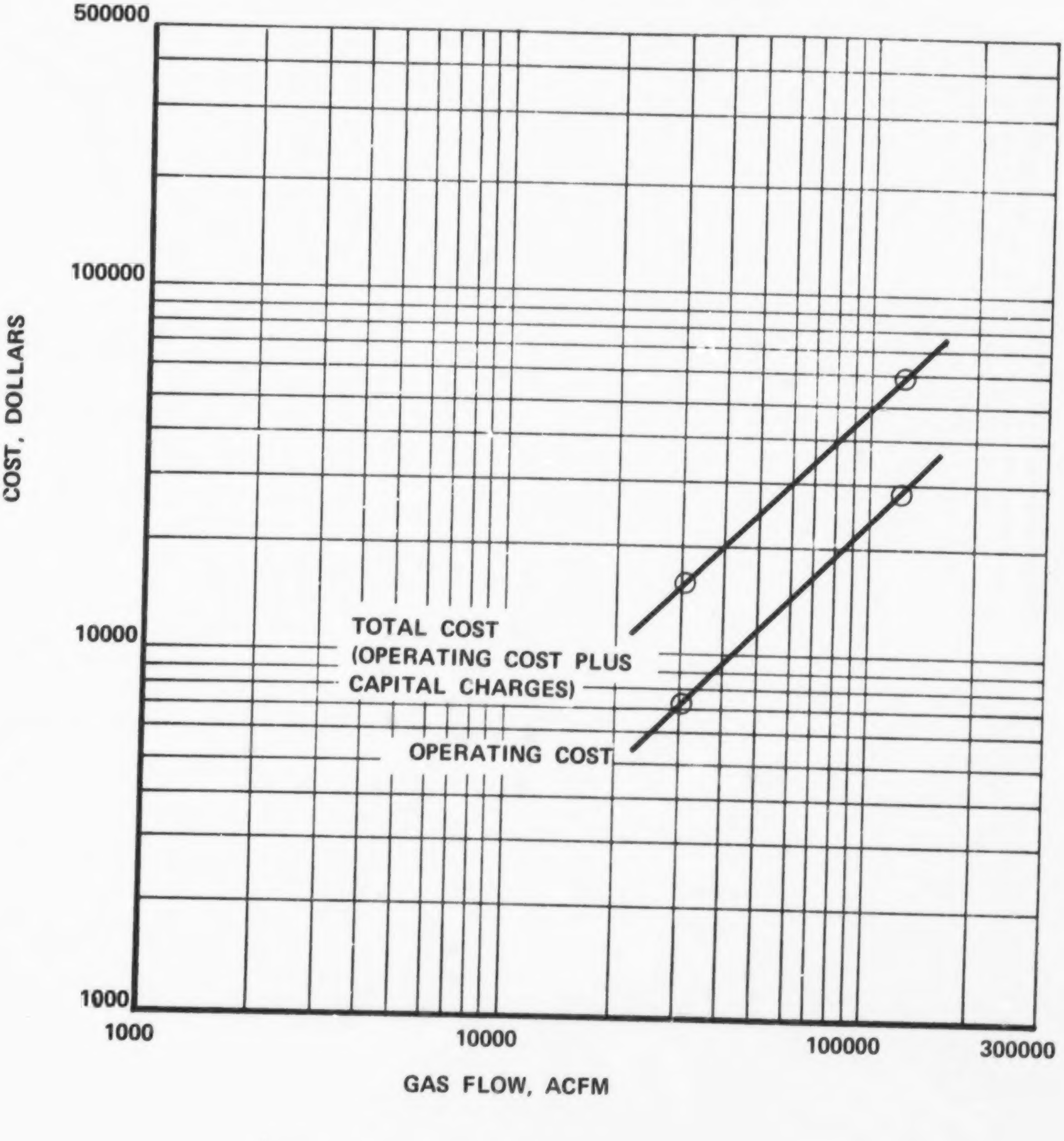


TABLE 170

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF FABRIC FILTERS FOR SOAP AND DETERGENT SPRAY DRYING

Population Size – 20

Sample Size – 2

Capital Cost = $35,277.5

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$ 24,022.1	$ 46,532.9
75	10,351.3	60,203.7

Capital Cost = $138,250

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$ 97,911.5	$178,589
75	48,916.7	227,583

FIGURE 134

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF FABRIC FILTERS FOR SOAP AND DETERGENT SPRAY DRYING

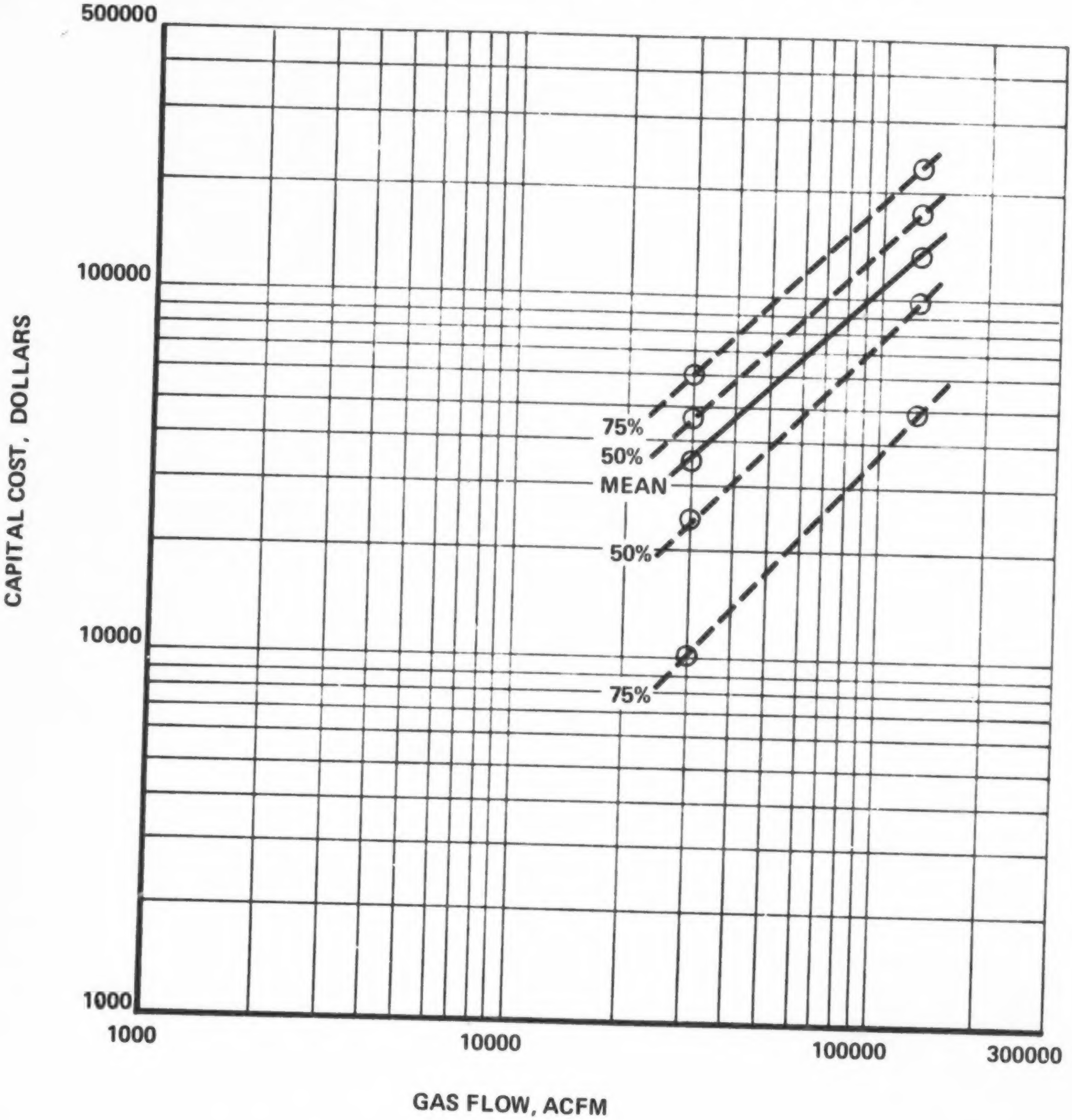


TABLE 171

FABRIC FILTER PROCESS DESCRIPTION FOR SOAP AND DETERGENT PRODUCT HANDLING SPECIFICATION

The fabric filter is to remove entrained detergent from the exhaust gas of the cyclone separator in the dry product pneumatic air conveying system. This device separates the air from the dried detergent beads being transferred from the spray dryer to dry product storage. Air used in the conveying system is atmospheric air. The fines recovered by the fabric filter represent an economically significant amount and are to be returned to the process.

The exhaust gas from the separating device will be brought to the filter by means of a fan. The fabric filter is to be located outside of the building in an area free of space limitations. The fan is to follow the filter.

The fabric filter is to consist of a minimum of five compartments and will allow for isolation of an individual compartment for cleaning during operation. Any four of these compartments should be capable of handling the full flow at designed differential pressure. Each section should also be capable of isolation for maintenance and have provisions for personal safety during filter operation. The collecting process is to be continuous and should include the following:

1. *Pulse jet fabric filter or equal, top removal design, compartmented as described above, with a maximum air-to-cloth ratio of 6/1 and a design differential pressure of 6 in. w.c.*
2. *Insulated hoppers with a minimum side angle of 60° and a manual slide gate at the screw conveyor for isolation. The hopper should be strong enough to support a full load, although in routine operation the hopper will not be full.*
3. *Insulate and heat trace entire system ductwork.*
4. *Carbon steel construction with manufacturer's standard prime paint applied.*
5. *An insulated screw conveyor equipped with a single rotary air lock.*
6. *A fan, located downstream of the filter sized with 20% excess capacity when operating at the design pressure and 90% of the maximum recommended speed.*

TABLE 172

FABRIC FILTER OPERATING CONDITIONS FOR SOAP AND DETERGENT PRODUCT HANDLING SPECIFICATION

Two sizes of fabric filters are specified for the "high efficiency" level. Vendors' quotations should consist of one quotation for each of the two sizes, with a representation of the efficiency expected for the unit quoted. The efficiency quoted may be better than the "high efficiency" case.

	Small	*Large*
Pneumatic Air Conveying System		
Product Rate, lb/hr	*20,000*	*80,000*
Inlet Gas to Fabric Filter		
Flow, ACFM	*20,000*	*80,000*
Temperature, °F	*100*	*100*
Flow, SCFM	*18,900*	*75,700*
% Moisture (vol)	*0.3*	*0.3*
Dew Point, °F	*15*	*15*
Inlet Solids Rate, lb/hr	*600*	*2,400*
Inlet Solids Loading, gr/ACF	*3.5*	*3.5*
Inlet Solids Loading, gr/DSCF	*3.7*	*3.7*
	High Efficiency	
Outlet Solids Rate, lb/hr	*1.71*	*6.86*
Outlet Solids Loading, gr/ACF	*0.01*	*0.01*
Outlet Solids Loading, gr/DSCF	*0.01*	*0.01*
Collection Efficiency, wt.%	*99.80*	*99.80*
Air-To-Cloth Ratio	*6/1*	*6/1*

TABLE 173

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR FABRIC FILTERS FOR SOAP AND DETERGENT PRODUCT HANDLING

			High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			20,000	80,000
°F			100	100
SCFM			18,900	75,700
Moisture Content, Vol. %			0.3	0.3
Effluent Contaminant Loading				
gr/ACF			3.5	3.5
lb/hr			600	2,400
Cleaned Gas Flow				
ACFM			20,000	80,000
°F			100	100
SCFM			18,900	75,700
Moisture Content, Vol. %			0.3	0.3
Cleaned Gas Contaminant Loading				
gr/ACF			0.01	0.01
lb/hr			1.71	6.86
Cleaning Efficiency, %			99.80	99.80
(1) Gas Cleaning Device Cost			24,768	98,993
(2) Auxiliaries Cost				
(a) Fan(s)				
(b) Pump(s)			8,665	24,338
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering			4,850	10,200
(b) Foundations & Support				
Ductwork				
Stack				
Electrical			24,025	91,400
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost			62,308	224,931

FIGURE 135

CAPITAL COSTS FOR FABRIC FILTERS
FOR SOAP AND DETERGENT PRODUCT HANDLING

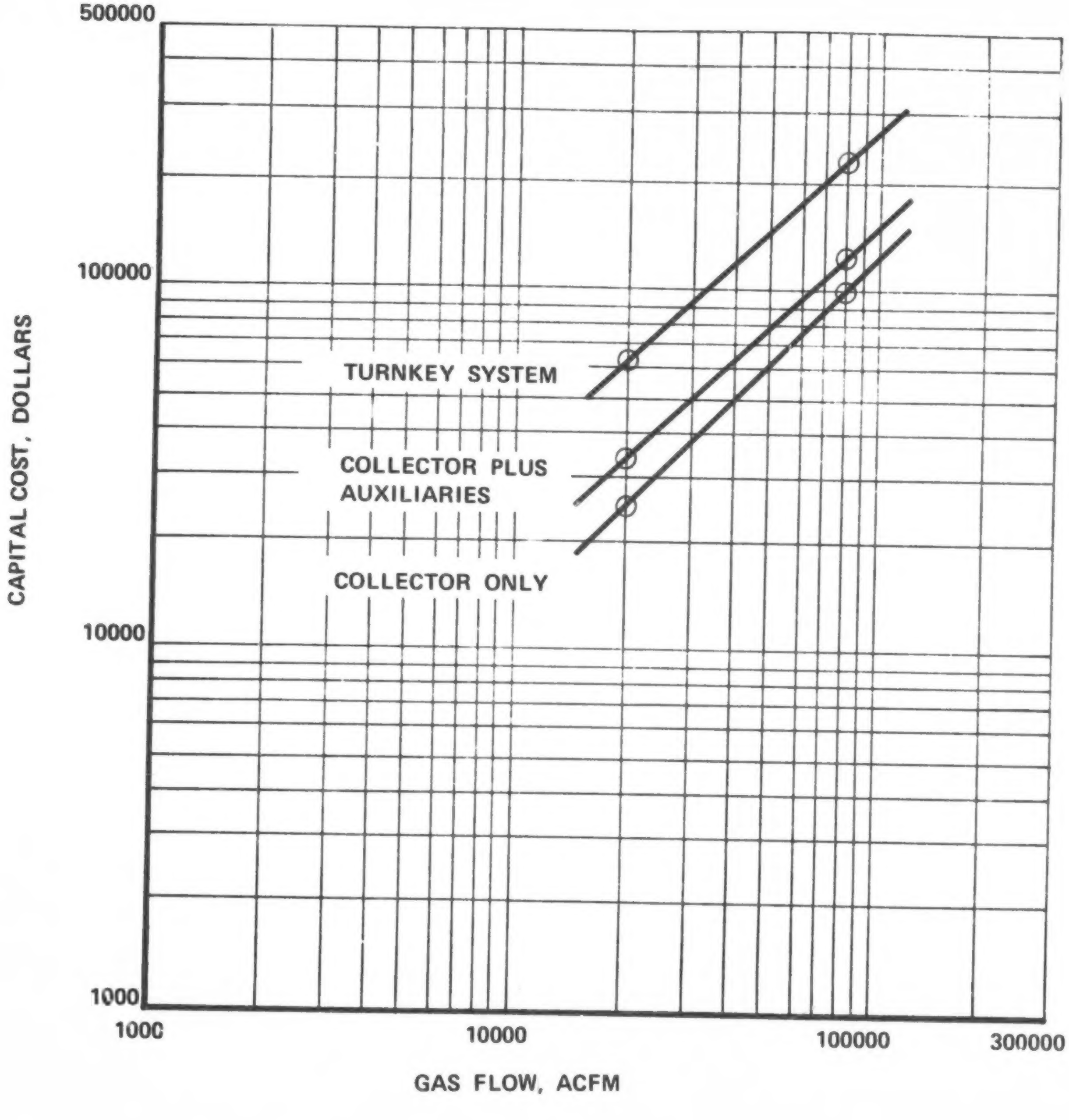


TABLE 174

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR FABRIC FILTERS FOR SOAP AND DETERGENT PRODUCT HANDLING

Operating Cost Item	Unit Cost			High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				8,000	
Operating Labor (if any)					
Operator	$8/hr			290	290
Supervisor					
Total Operating Labor				290	290
Maintenance					
Labor	$6/hr			975	3,250
Materials				493	1,740
Total Maintenance				1,468	4,990
Replacement Parts				775	2,700
Total Replacement Parts				775	2,700
Utilities					
Electric Power	$.011/kw-hr			2,841	11,704
Fuel					
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utiiities				2,841	11,704
Total Direct Cost				5,374	19,684
Annualized Capital Charges				6,231	22,493
Total Annual Cost				11,605	42,177

FIGURE 136

ANNUAL COSTS FOR FABRIC FILTERS FOR SOAP AND DETERGENT PRODUCT HANDLING

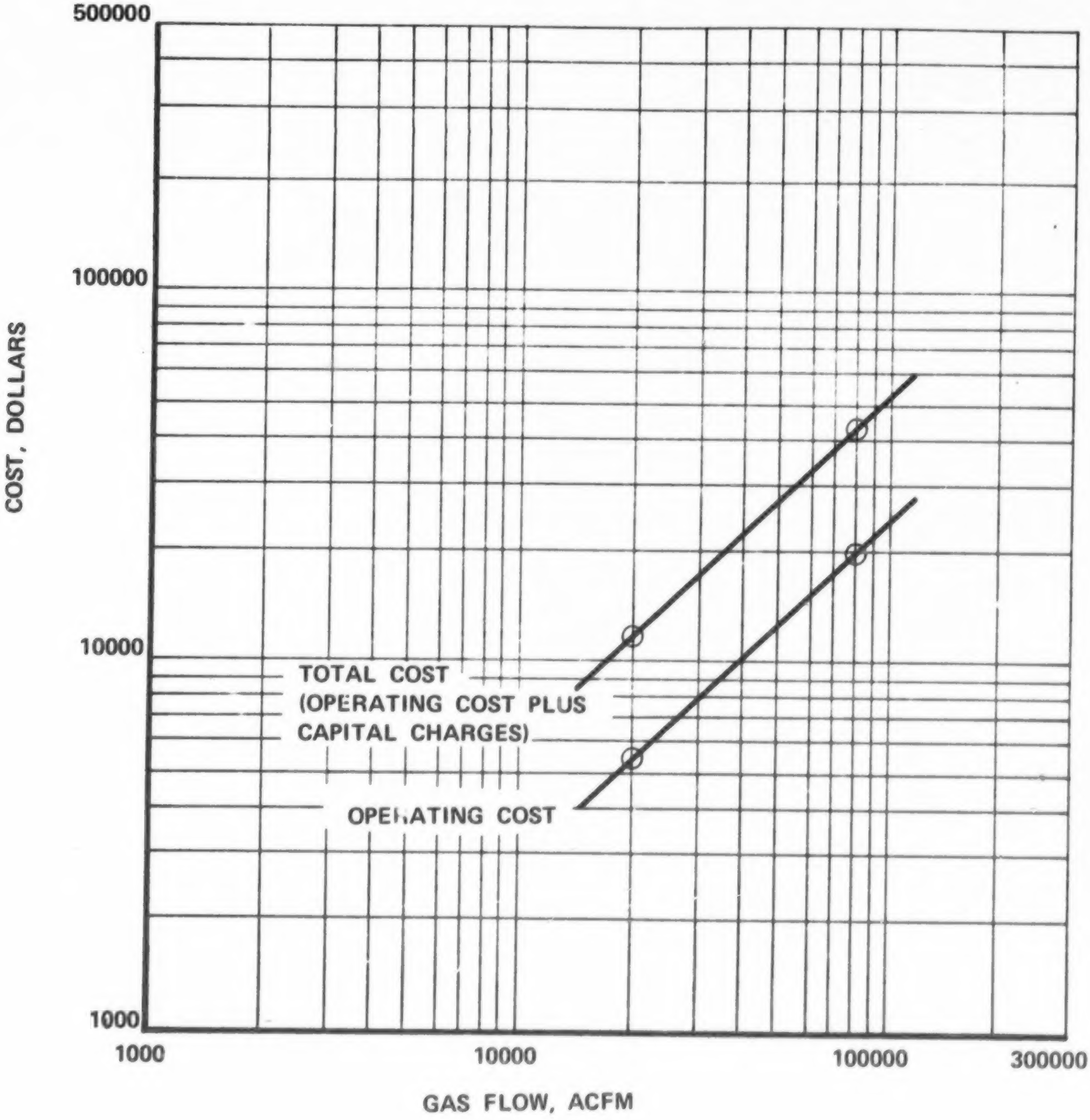


TABLE 175

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF FABRIC FILTERS FOR SOAP AND DETERGENT PRODUCT HANDLING

Population Size – 20　　Sample Size – 2

Capital Cost = $24,768.5

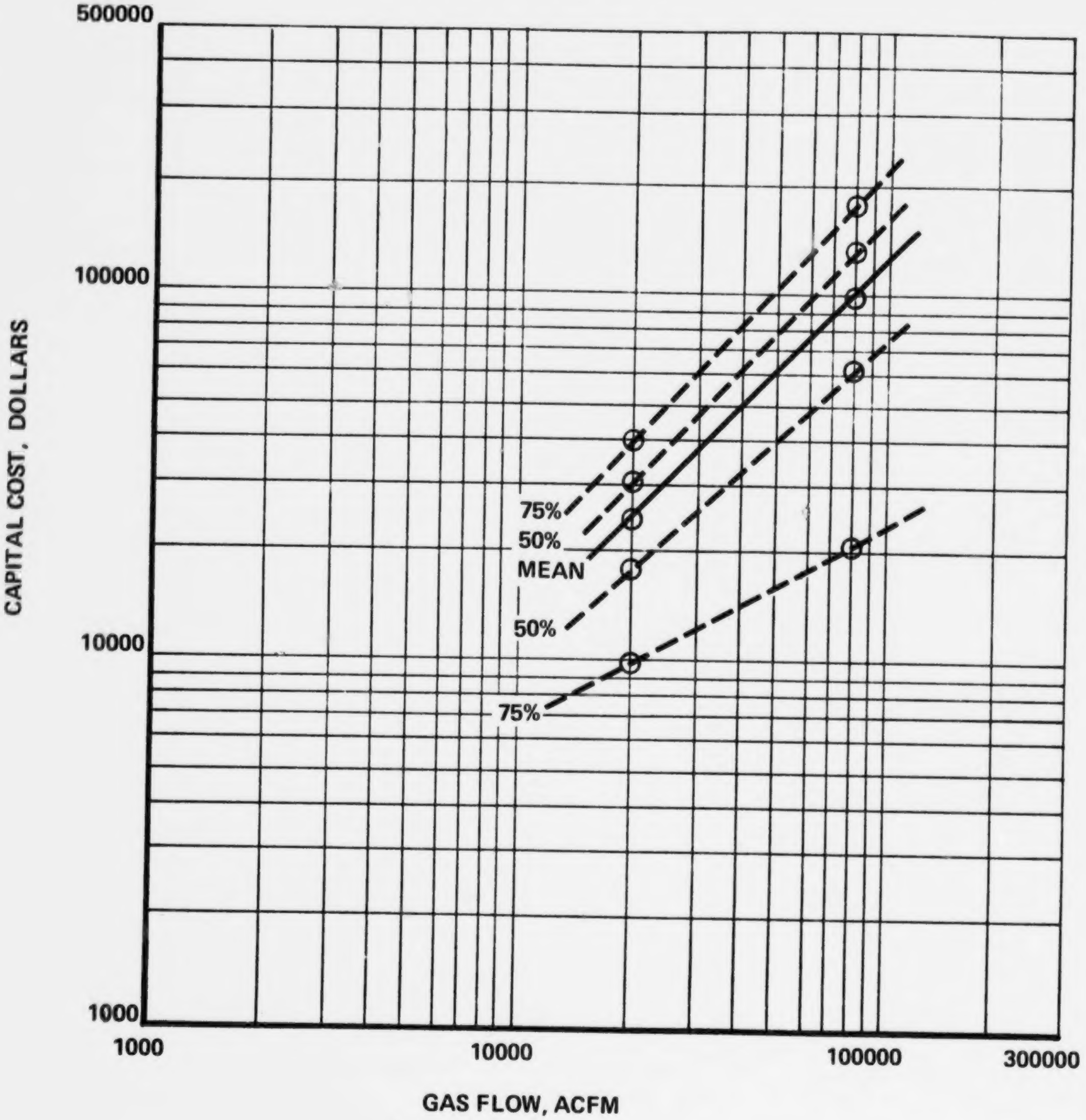
Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$ 18,007.6	$ 31,529.4
75	9,795.99	39,741.0

Capital Cost = $98,993.5

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$ 63,815.5	$134,171
75	21,088.8	176,898

FIGURE 137

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF FABRIC FILTERS FOR SOAP AND DETERGENT PRODUCT HANDLING



REFERENCES

1. "World Soap, Detergent Production," Soap & Chem. Specialties, 46: p. 40-4+; (Sept. 1970)

2. Duckworth, R. A., "Synthetic Detergent Powders," Chem. and Proc. Engineering, 51:4, p. 63-65+ (1970)

3. Silvis, S. T.; "Detergents," Chemical Week, 105:12; p. 72-4+, (1969)

4. Hatch, L. F., "What's Ahead for Synthetic Detergents," Hydrocarbon Processing, 47, 3, p. 93 (1968)

5. Marquis, D. M., "Make AOS from Olefins and SO_3," Hydrocarbon Processing, 47, 3, p. 109-114 (1968)

6. Hollis, G. L., "Domestic Detergents: today-tomorrow," Chemistry and Industry; p. 1309-15; October 10, 1970

7. Kremers, F. J., and Shultz, A., "Continuous Sulfonation with Sulfur Trioxide," Soap and Chem. Specialties, 46: p. 44-6+ (1970)

8. Holzman, S. and Zoller, U., "Long Chain Alcohol Sulfates from Converter Gas," Soap and Chem. Specialties, 45, 9, p. 43-46+ (1969)

9. Gilbert, E. E., Sulfonation and Related Reactions, Interscience Publishers, N.Y., 1965

10. Shreve, R. N., The Chemical Process Industries, McGraw-Hill Book Company, Inc. N.Y., 1956

11. Danielson, John A., Air Pollution Engineering Manual, Public Health Service, Cincinnati, 1967, p. 716-20.

12. Spitz, L., "Advances in Saponification, Drying and Soap Finishing Technology," Soap and Chem. Specialties 44: p. 60+ (October, 1968)

13. Phelps, A. H., "Air Pollution Aspects of Soap and Detergent Manufacture," Journal of Air Pollution Control Association, 17, 8, p. 505-7 (1967)

14. "Market Newsletter," Chemical Week, p. 33, October 18, 1972

6. LIME KILNS

The manufacture of lime is an ancient practice, dating back to the Roman, Greek, and Egyptian civilizations. In the colonial period of America, lime was manufactured by burning limestone in brick or masonry kilns dug out of the side of a hill. The kilns used either a coal or wood fire at the bottom, and were open at the top. Lime manufacture did not change substantially until the twentieth century, and most of the progress has been made since 1935.

Today, the manufacture of lime is a large industry using sophisticated technical control and making a product of high quality and uniformity. The manufacture of lime involves the burning of limestone ($CaCO_3$ or $CaCO_3 \cdot MgCO_3$) to release carbon dioxide and form quicklime (CaO or $CaO \cdot MgO$). There are two types of limestone used to produce lime. The limestone is classified as "high calcium" or "calcite" if the magnesium carbonate content is less than 5%, and "dolomitic limestone" or "dolomite" if the magnesium carbonate content is 30 to 40%.(1) Figure 88 is a flowsheet showing the manufacture of lime in its entirety.(2)

In the United States, limestone is calcined in either vertical or rotary kilns. In 1964, about 15% of the open-market lime and over 50% of the captive market lime was produced in vertical kilns.(3) In that year about 30% of the 11.4 million tons of lime produced was captive market lime.(2) This "captive market" lime consists mainly of the lime produced and used in the alkali, sugar, and pulp and paper industries.

This section will discuss vertical limestone calcining kilns and those rotary lime kilns used in the pulp and paper industry. Other rotary kilns were studied under NAPCA Contract CPA70-150.

VERTICAL KILNS

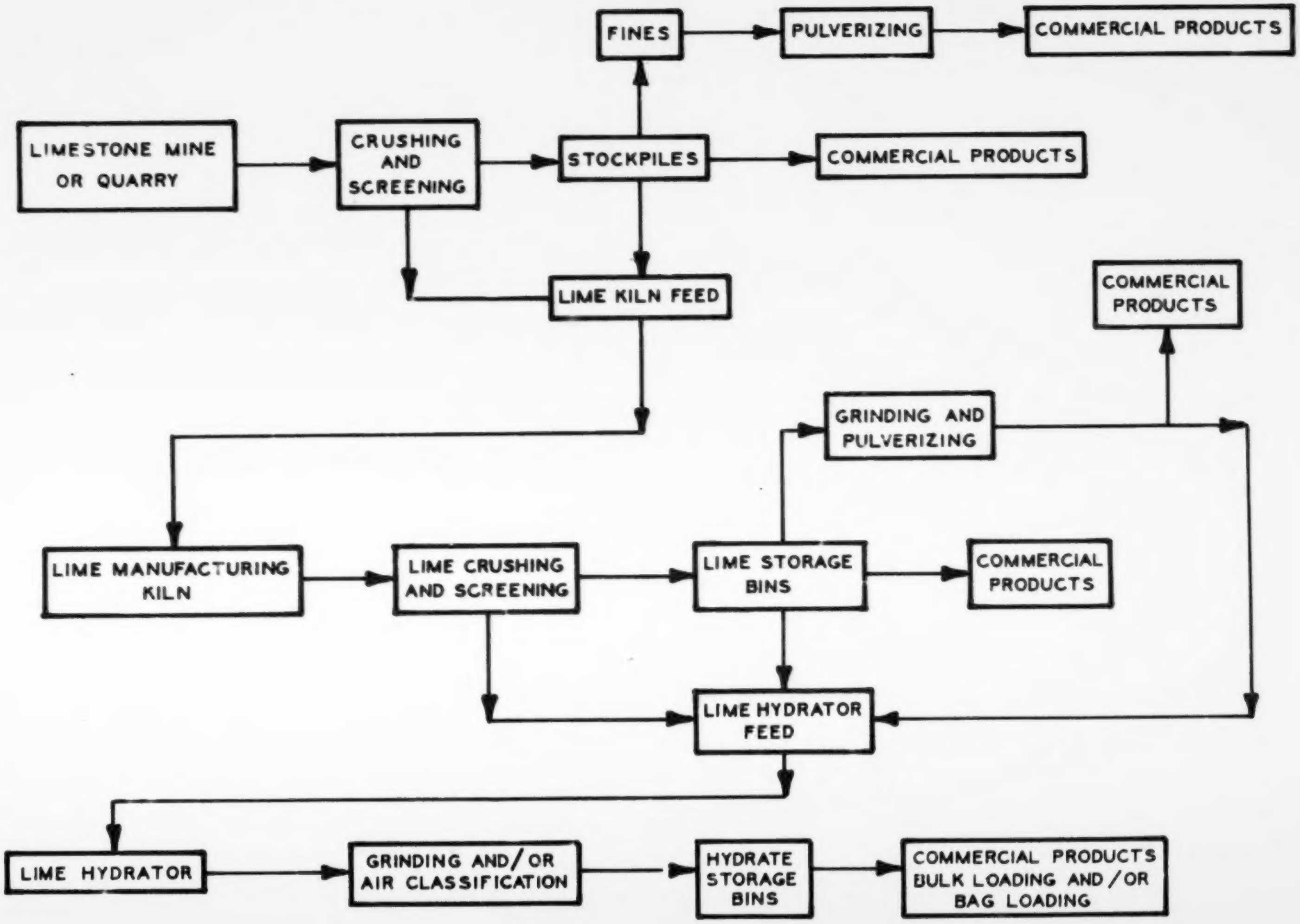
Process Description

Vertical, or shaft, kilns are the oldest and most widely used type of lime calcining kilns. Consequently, the particulars of kiln design vary greatly. All vertical kilns do operate similarly, however, and have four distinct zones from top to bottom:

1. Stone storage zone
2. Preheating zone
3. Calcining zone
4. Cooling and discharge zone

FIGURE 138

SIMPLIFIED FLOW SHEET OF A TYPICAL LIME MANUFACTURING OPERATION


FINES
PULVERIZING
COMMERCIAL PRODUCTS
LIMESTONE MINE OR QUARRY
CRUSHING AND SCREENING
STOCKPILES
COMMERCIAL PRODUCTS
LIME KILN FEED
COMMERCIAL PRODUCTS
GRINDING AND PULVERIZING
LIME MANUFACTURING KILN
LIME CRUSHING AND SCREENING
LIME STORAGE BINS
COMMERCIAL PRODUCTS
LIME HYDRATOR FEED
LIME HYDRATOR
GRINDING AND/OR AIR CLASSIFICATION
HYDRATE STORAGE BINS
COMMERCIAL PRODUCTS BULK LOADING AND/OR BAG LOADING

These zones are shown in a schematic diagram of a vertical kiln contained in Figure 139.(2,3)

The flow of stone in the kiln is countercurrent to the flow of cooling air and combustion gases. The stone is charged at the top and preheated by the hot exhaust gases from the calcining zone. The stone is burned in the calcining zone in various ways which will be described below. Air blown into the bottom of the kiln cools the lime before it is discharged. This air is heated sufficiently by the time it reaches the calcining zone to be used as secondary combustion air. The lime is discharged to cars on tracks or to conveyor belts, and either shipped or further processed by hydrating.

The limestone is usually charged to the kiln continuously by inclined conveyor belt, although some older kilns still are charged in batches by skip hoist. Modern vertical kilns are enclosed and vent exhaust gases through flues or dampers. In these kilns either a forced or induced draft moves the gases through the kiln. There are some old kilns left that are open at the top, but they have serious disadvantages. Winds interfere with the draft, sometimes causing downdrafts, or rains wet the stone charge.

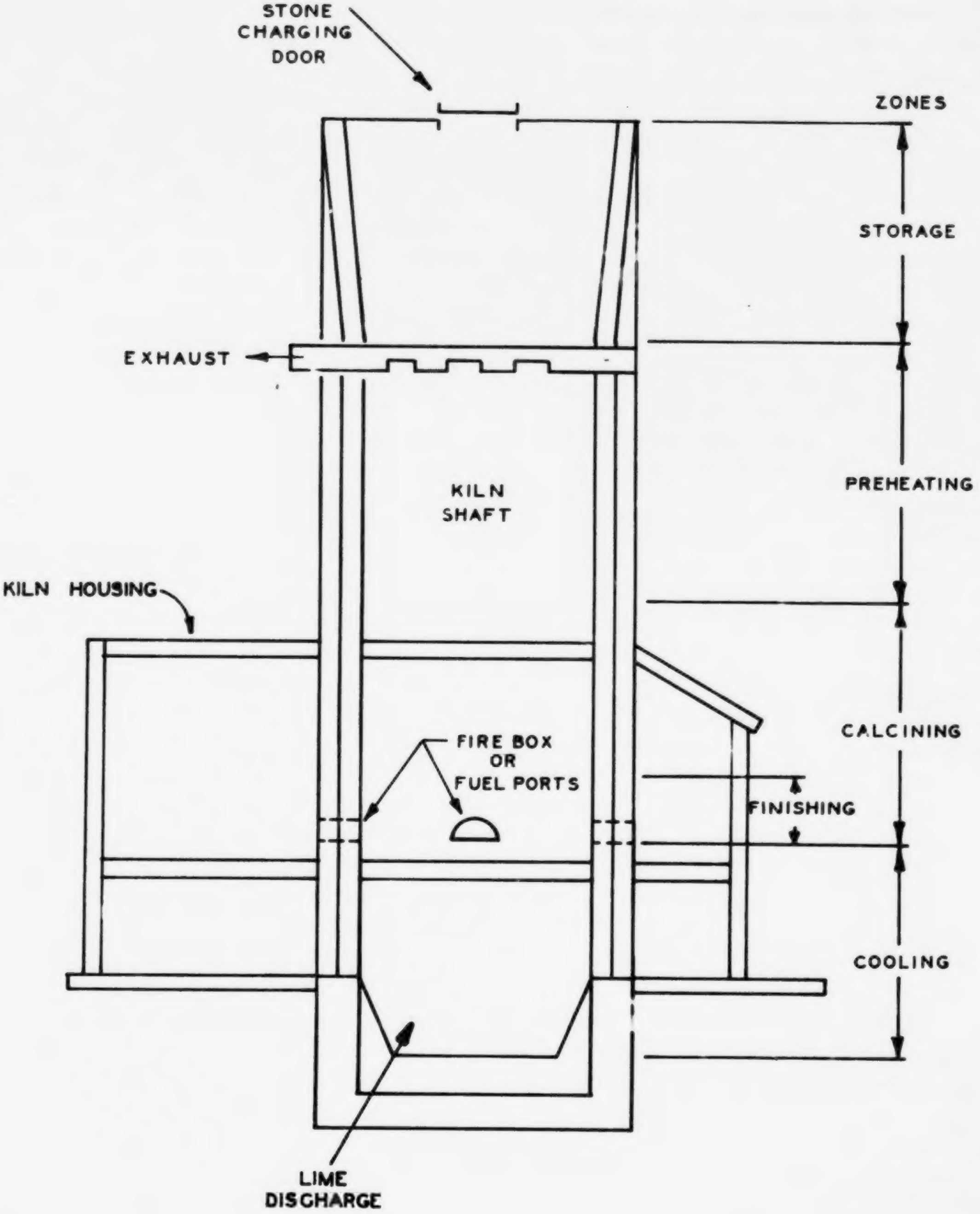
The hot exhaust gases rising from the calcining zone contain more than enough heat for preheating, and the excess is frequently recycled to the calcining zone and mixed with cooling air to be used as secondary combustion gas. This exhaust gas recirculation is one reason that emission rates from enclosed vertical kilns are generally lower than those of rotary kilns.

In most modern lime kilns, gaseous rather than solid fuels are employed. This prevents a reduction in quality due to fuel ash. The gaseous fuels employed are natural gas, fuel oil, or producer gas. Natural gas is preferred in the United States. Most modern natural gas and oil fired kilns are equipped with center burners on two or four sides of the kiln in the calcining zone. Side burners are usually used also. This arrangement provides good uniformity in temperature throughout the calcining zone, improving kiln performance, fuel efficiency, lime quality, and kiln capacity. Producer gas kilns burn solid fuel in separate furnaces. The fuel is burned with a deficient air supply, or with an air-steam mixture, producing carbon monoxide and hydrogen. This hot combustible gas is burned in the calcining zone, either undiluted or diluted with primary air.

There are two types of kilns where solid fuel is in direct contact with the limestone. The standard vertical kiln, which was common in the United States up to 1940, is largely of the direct hand fired type. These kilns are manually

FIGURE 139

VERTICAL LIME KILN


STONE CHARGING DOOR
ZONES
STORAGE
EXHAUST
PREHEATING
KILN SHAFT
KILN HOUSING
CALCINING
FIRE BOX OR FUEL PORTS
FINISHING
COOLING
LIME DISCHARGE

fueled by periodic coal shoveling. Usage of kilns of this type is declining because of considerable manual labor, low kiln capacity, poor thermal efficiency, and inadequate quality and uniformity of lime.(3) The second type of solid fuel kiln is the mixed feed kiln. This kiln is widely used in Europe although it produces a generally poorer grade of lime. The fuel used in mixed feed kilns is usually anthracite coal or a metallurgical grade of low reactivity coke. The fuel is comparable in size with the limestone and the two are intimately mixed in fixed proportions before charging. The overriding advantage of a mixed feed kiln is its fuel economy. It has the lowest fuel consumption of any type of lime kiln, and is a great improvement over earlier solid fuel kilns. In these older solid fuel kilns, substantial heat losses occur as the result of carbon monoxide formation. Carbon dioxide reacts with hot fuel to form carbon monoxide with a subsequent loss of heat ($CO_2 + C + \text{heat} \rightarrow 2CO$). For 1% CO in the exhaust gas, the heat loss will be about 240,000 Btu/ton of lime.(3) The intimate fuel-limestone mixing in mixed feed kilns greatly reduces the possibility of fuel-carbon dioxide contact and the occurrence of the resultant heat loss.

Vertical kilns range from 6.5 to 25 ft in outside diameter and 16 to 100 ft in height; an average size is 12 ft in diameter by 50 ft in height. Lime kilns are generally constructed of insulated refractory brick, and a heavy steel boiler plate casing. Kiln capacities vary greatly; the older standard kilns have capacities from 1/4 to 1/2 ton/day/ft^2 cross-sectional area, while large direct gas fired kilns have 1 to 2 ton/day/ft^2 capacities. The majority of kilns have capacities ranging from 50 to 250 plus tons of lime/day.(3) Some kilns recently built have capacities up to 680 ton/day.(4) Fuel consumption for direct and indirect gas fired kilns range from 4 to 6 million Btu/ton of lime produced. For large mixed feed kilns, consumption of 3 million Btu/ton has been reported.(3)

The vertical kiln has several advantages over a rotary kiln:

1. Lower fuel consumption.
2. Greater flexibility due to smaller producing units.
3. Lower first cost.

The major disadvantages are:

1. Cannot effectively handle small feed sizes.
2. Chemical reactivity is less, producing less product uniformity.
3. Longer calcining time required.
4. Lesser degree of quality control possible.

Process Chemistry

The calcination of limestone involves thermal decomposition and can be represented by the following reactions:

1. $CaCO_3(c) + \text{Heat} \rightleftarrows CaO(c) + CO_2(g)$
   (high calcium limestone) (high calcium quicklime)
2. $CaCO_3 \cdot MgCO_3(c) + \text{Heat} \rightleftarrows CaO \cdot MgO(c) + 2CO_2(g)$
   (dolomitic limestone) (dolomitic quicklime)

There are three essential factors in limestone decomposition kinetics:(3)

1. The limestone must be heated to the dissociation temperature of carbonates. This temperature is 1648° F for calcite and from 930 to 1480° F for dolomite. The dissociation temperature of dolomite is dependent on the amount of magnesium carbonate ($MgCO_3$) present, as it dissociates at lower temperatures (756 to 896° F). The heat required to reach the theoretical minimum dissociation temperature is approximately 1.5 million Btu/ton of high calcium quicklime and 1.25 million Btu/ton of dolomitic quicklime produced.

2. The minimum dissociation temperature must be maintained throughout the reaction. The heat consumed here is about 2.77 million Btu/ton of high calcium quicklime and 2.6 million Btu/ton of dolomitic quicklime produced. In practice, however, higher temperatures are maintained and more heat is consumed. Inadequate control of the temperature will lead to fuel losses and poor product quality. The correct temperature depends on the limestone being calcined, but generally ranges from 1900 to 2450° F for calcium carbonate and 1750 to 2300° F for dolomitic limestone.

3. The CO_2 gas that is produced must be removed to minimize carbonate reformation, as the reaction is reversible.

Raw Materials and Products

The feed to lime kilns consists of the carbonates of calcium or magnesium, which are obtained from natural deposits of limestone, marble, chalk, dolomite, or oyster and clam shells. Limestone is generally used as the raw material for lime manufacture. Some plants, particularly along the Gulf Coast, use oyster shells. Limestone deposits exist throughout the country, but only a small portion are of sufficient purity to be used in lime manufacture. Table 176(2) presents some typical lime kiln feed compositions.

Greater than 90% of the limestone quarried is from open pit operations. The remainder is extracted from underground mines.(2) The stone that is quarried contains as few impurities as possible. These impurities include silica, alumina, iron and clay.

The size distribution used in vertical lime kilns is critical. Ample voids must be present in the bulk limestone to insure unrestricted circulation of flue gases and air around and through the stone and lime. This is necessary for uniform heat transfer and expulsion of the CO_2. The ideal external void fraction is 45%.(3) A range of particle sizes may have lower void fraction (i.e. the small particles fill in between the large ones) than either uniform large particles or uniform small particles. Stones less than 2 inches in diameter cannot be successfully calcined in a vertical kiln. Stone sizes range from 8 by 12 in. to a minimum of 2 by 4 in., with an average size being 4 by 8 in.(3)

Limestone and lime have as many different uses in industry as any natural substance. Lime is a close second to sulfuric acid in total production among the basic chemicals. Lime is used in building construction, agriculture, and highway construction. Lime is used in the pulp and paper industry to causticize the waste sodium carbonate solution in the sulfate process. This particular process will be discussed in detail in the second half of this section. Lime is also used in the manufacture of pig iron, steel, magnesia or magnesium metal, alkalis, sugar, calcium carbide, insecticides, bleaches, and other chemicals. Lime, in terms of tonnage, is the main water treatment chemical. It is employed in water softening, purification, coagulation, neutralization of acid wastes, sewage treatment, and treatment of industrial waste water. Lime has also been used in the form of a slurry for the scrubbing of stack gases to remove SO_2, HCl, HF, and other acidic gases.

Most lime produced is high calcium quicklime of at least 90% calcium oxide content. Table 177 shows a typical composition of high calcium and dolomitic quicklime.(3) The composition of the lime varies depending on its intended usage. This composition is controlled by proper selection of the stone and specific conditions of the manufacturing process used.

ROTARY KILNS (PULP MILL)

Rotary lime sludge kilns are widely used in alkali pulping processes in the

TABLE 176

TYPICAL ANALYSES OF HIGH CALCIUM AND DOLOMITIC COMMERCIAL LIMESTONES(2)

Component	High Calcium Wt.%	Dolomitic Wt.%
Calcium carbonate ($CaCO_3$)	97.40	52.34
Magnesium carbonate ($MgCO_3$)	1.25	47.04
Iron oxide (Fe_2O_3)	0.11	0.04
Aluminum oxide (Al_2O_3)	0.35	0.20
Silica (SiO_2) plus acid insolubles	0.95	0.26
Loss on ignition (CO_2)	43.40	47.67

TABLE 177

TYPICAL COMPOSITION OF HIGH CALCIUM AND DOLOMITIC QUICKLIME(3)

Component	High Calcium, Range, wt.%	Dolomitic, Range, wt.%
CaO	93.00 to 98.00	55.00 to 57.50
MgO	0.30 to 2.50	37.00 to 41.00
SiO_2	0.20 to 2.00	0.10 to 1.75
Fe_2O_3	0.10 to 0.40	0.05 to 0.40
Al_2O_3	0.10 to 0.60	0.05 to 0.50
CO_2	0.40 to 2.00	0.40 to 2.00
SO_3	0.01 to 0.10	0.01 to 0.10
P	trace to 0.05	trace to 0.05

pulp and paper industry. These processes basically convert wood to wood pulp by "cooking" the wood in an alkaline solution known as the cooking liquor. The function of the kiln is to reburn lime for use in the regeneration of this wood cooking liquor.

Alkali pulping is a term applicable to two similar processes used in the preparation of cellulose wood pulp. These are the sulfate pulping (Kraft) and the soda pulping processes. In both processes, sodium hydroxide is the major ingredient in the cooking liquor. The main difference between the processes is that sodium sulfate is used in the former instead of sodium carbonate as a make-up chemical for the cooking liquor. The sulfate is reduced to sulfide, so in the sulfate process, not only caustic soda but also sodium sulfide is a component of the cooking liquor. A simplified sulfate pulping flow diagram is shown in Figure 140.(5,6)

Process Description

In sulfate pulping, wood chips are cooked in a digester at elevated temperature and pressure. The "fresh" cooking liquor is a solution of NaOH, Na_2S, and Na_2CO_3, called the "white liquor." The spent liquor from the digester is separated from the pulp. This liquor, known as "black liquor," contains 95 to 99% of the alkali, mostly in the form of sodium carbonate. Various organic sulfur compounds in combination with sodium sulfide are also present, along with other numerous impurities. A portion of this "black liquor" is recycled back to the digester. The actual cooking liquor will contain from 20 to 50% of this recycled liquor. The remainder of the black liquor is recovered and regenerated.

The recovery of the cooking chemicals involves the concentration of the black liquor by evaporation, burning of the concentrated liquor in a furnace to convert the sodium compounds to Na_2CO_3 and Na_2S, and causticizing the carbonate "smelt" with lime to reform the white liquor. The lime used in causticizing is regenerated by burning the $CaCO_3$ mud, which is generated in causticizing, in a rotary lime kiln. In this way, about 90 to 95% of the total sodium and 50% of the total sulfur are recovered.(7) Since the function of the lime kiln is an integral part of the black liquor recovery process, the total process will be described briefly. A flow chart of the process is shown in Figure 141.

The molten smelt from the recovery furnace is removed and dissolved in water and weak liquor from the causticizer, forming "green liquor." This liquor

FIGURE 140

SIMPLIFIED KRAFT MILL FLOW DIAGRAM

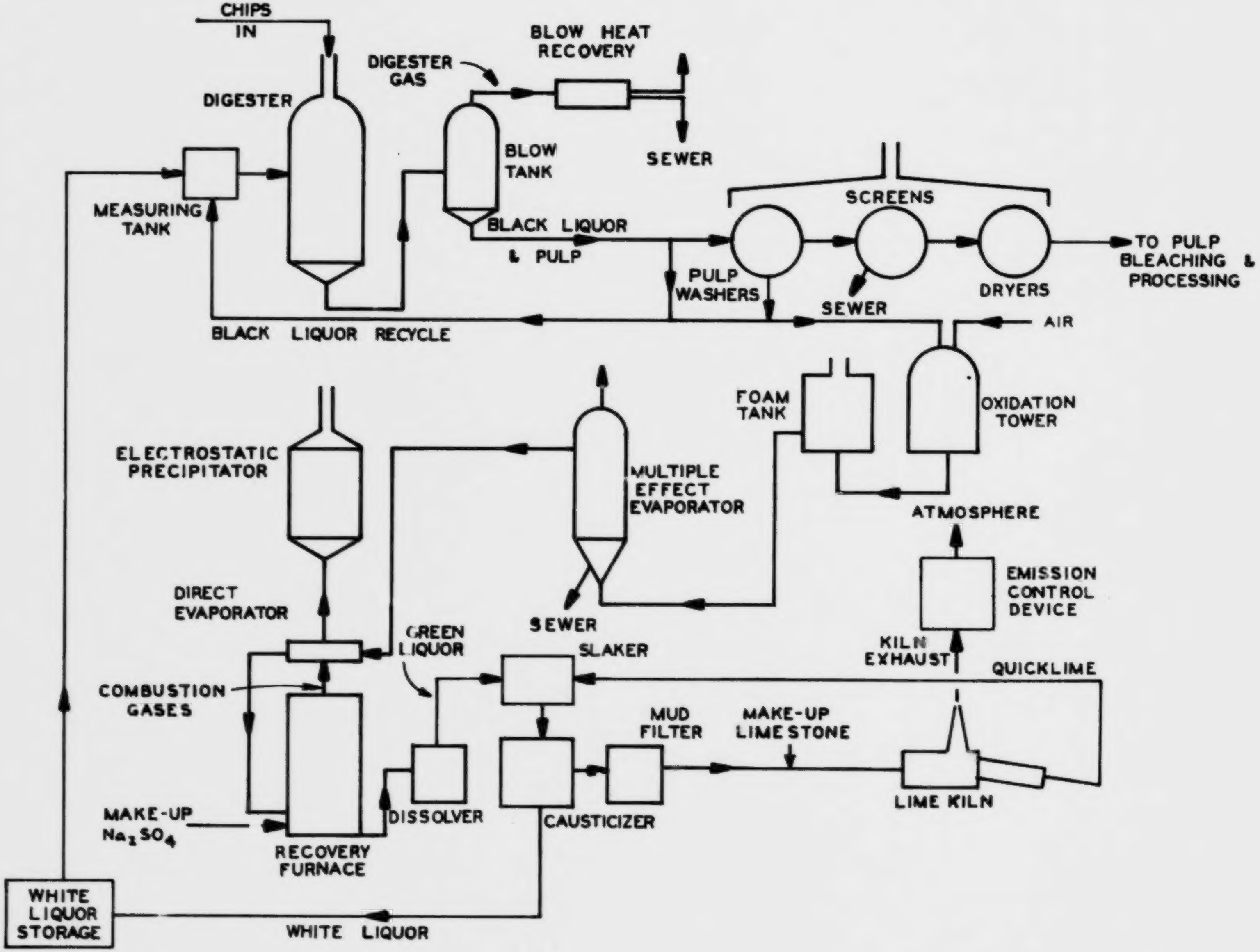
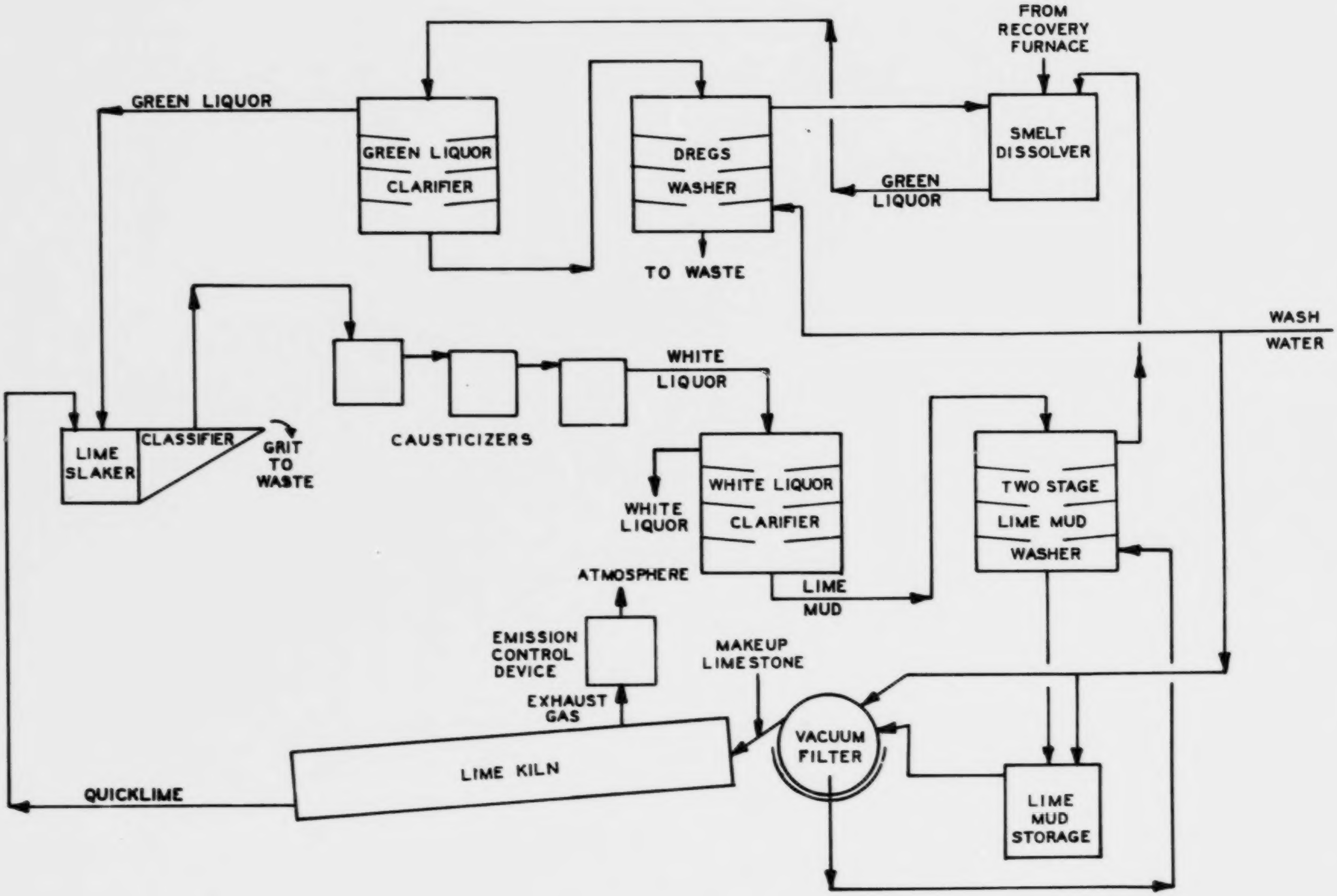

CHIPS IN
DIGESTER
MEASURING TANK
DIGESTER GAS
BLOW HEAT RECOVERY
BLOW TANK
SEWER
BLACK LIQUOR & PULP
SCREENS
PULP WASHERS
SEWER
DRYERS
TO PULP BLEACHING & PROCESSING
BLACK LIQUOR RECYCLE
AIR
FOAM TANK
OXIDATION TOWER
ELECTROSTATIC PRECIPITATOR
MULTIPLE EFFECT EVAPORATOR
ATMOSPHERE
EMISSION CONTROL DEVICE
DIRECT EVAPORATOR
SEWER
GREEN LIQUOR
SLAKER
KILN EXHAUST
QUICKLIME
COMBUSTION GASES
MUD FILTER
MAKE-UP LIMESTONE
LIME KILN
MAKE-UP Na2SO4
DISSOLVER
CAUSTICIZER
RECOVERY FURNACE
WHITE LIQUOR STORAGE
WHITE LIQUOR

FIGURE 141

KRAFT PULPING RECAUSTICIZING FLOWSHEET


FROM RECOVERY FURNACE
GREEN LIQUOR
GREEN LIQUOR CLARIFIER
DREGS WASHER
SMELT DISSOLVER
GREEN LIQUOR
TO WASTE
WASH WATER
WHITE LIQUOR
LIME SLAKER
CLASSIFIER
GRIT TO WASTE
CAUSTICIZERS
WHITE LIQUOR
WHITE LIQUOR CLARIFIER
TWO STAGE LIME MUD WASHER
ATMOSPHERE
LIME MUD
EMISSION CONTROL DEVICE
MAKEUP LIMESTONE
EXHAUST GAS
VACUUM FILTER
LIME KILN
QUICKLIME
LIME MUD STORAGE

will contain 20% Na_2CO_3 and 5% Na_2S by weight, expressed as Na_2O.(8) The green liquor is clarified to remove such impurities as Fe_2O_3, silica, alumina, unburned carbon, and furnace brick. Quicklime from the lime kiln is converted to slaked lime by mixing it with clarified green liquor. This mixture is sent to a series of agitated tanks known as causticizers. In these vessels, calcium hydroxide reacts with sodium carbonate to form sodium hydroxide and calcium carbonate. The sodium sulfide does not react in this mixture. The lime sludge is separated from the clean white liquor by a clarifier. The lime sludge mainly consists of calcium carbonate and sodium hydroxide. This lime mud is washed countercurrently, finally with fresh water on a rotary vacuum filter. The sodium content should be reduced by washing from about 22 wt.% to about 0.5 wt.% dry basis, but the lime mud may have about 1.5 to 3.0 wt.% sodium (in the form of Na_2O) due to inadequate washing.(8)

The soda pulping process operates in a similar way with the exception that sodium carbonate is used in place of sodium sulfate and, therefore, no sulfides or sulfates are present at any stage of the operation.

The lime sludge from the vacuum filter, which contains 55 to 60% solids, is the feed material for the rotary lime kiln. Some fresh limestone is usually also burned to replenish lost lime. The kiln, which is supported by rollers, is a long inclined horizontal cylinder made from refractory brick and lined with steel plates. Most kilns rotate at a speed of about 1 rpm. These kilns range in size from 6 to 11 ft in diameter and 100 to 325 ft in length.(9) Kilns in these ranges produce from 30 to 175 tons of lime per day. Kilns built today are a bit larger. The lime flows countercurrent to the heat, the lime mud entering at one end and the hot air entering at the other. Most kilns are fired by either oil or natural gas. The kiln is composed of three fairly distinct zones:(7)

1. The feed and drying zone.
2. The central or preheating zone.
3. The calcining zone.

In the drying zone, steel chains are commonly used to facilitate heat transfer from the hot gases to the lime mud. As the kiln rotates, these chains will alternatively contact the hot gases and the mud. The temperature in this end of the kiln is kept below 1,000° F. In the preheating zone, the lime will pass through a plastic or semi-liquid state, forming pellets up to 1 in. in diameter under normal conditions.(8) The operating temperature in the preheating and calcining zones is generally in the 2,000 to 2,400° F range, with higher temperatures being found in shorter kilns. The quicklime is discharged directly to the slaker or to a surge bin and then to the slaker. A typical composition of this reburned quicklime is shown in Table 178.(10)

TABLE 178

TYPICAL COMPOSITION OF REBURNED QUICKLIME(10)

Component	Wt.%
CaO	93
Na_2O	2
MgO	1
Al_2O_3	1
Fe_2O_3	1
SiO_2	1
CO_2	1
Total oxides	100

Process Chemistry

The process chemistry of rotary lime sludge kilns is generally the same as vertical kilns. The heat requirement is greater because the lime mud contains 40 to 45% water. The heat requirement for lime with 40% water is around 6,740,000 Btu/ton of lime.(7)

Several reactions take place in the different stages of the black liquor recovery process.

1. Recovery furnace
   a. Various organic combustion reactions.
   b. $Na_2O + CO_2 \rightarrow Na_2CO_3$
   c. $Na_2SO_4 + 2C \rightarrow Na_2S + 2CO_2$
2. Slaker
   a. $CaO + H_2O \rightarrow Ca(OH)_2$
   b. $Ca(OH)_2 + \underset{\text{(small extent)}}{Na_2CO_3} \rightleftarrows 2NaOH + CaCO_3 \downarrow$
3. Causticizer
   a. $Ca(OH)_2 + Na_2CO_3 \rightleftarrows 2NaOH + CaCO_3 \downarrow$
4. Lime Kiln
   a. $CaCO_3 \rightarrow CaO + CO_2$

NATURE OF THE GASEOUS DISCHARGE

The air pollutants emitted from both types of lime kilns consist mainly of lime dust and the products of combustion. Emissions from a lime sludge kiln will also include a small percentage of sodium and sulfur compounds. As a result of this, odor can be a problem of lime sludge kiln exhausts.

The contaminants in vertical lime kiln gases will be similar to those found in rotary lime kiln exhaust gases. The emissions from a vertical kiln will present less of a problem for the following reasons:

1. A portion of the exhaust gas from a vertical kiln is usually recycled to salvage its heat content.

2. The particulate emissions from a vertical kiln are not as small as in the rotary kiln case because there is much less attrition in the former than in the latter.

3. The temperature of the exhaust gas from a vertical kiln is in the 225 to 310° F range, as opposed to a rotary kiln exhaust temperature of 300 to 1,200° F.

The exhaust gases principally consist of carbon dioxide, water, and nitrogen. Depending on the type of fuel used, various other gaseous contaminants will be present. Typical exhaust compositions from a gas fired lime sludge kiln are presented in Table 179.(12) The relative usage of fuels for limestone calcining is given in Table 180.(3) Oil and natural gas are predominantly used in lime sludge kilns.

If sulfur-containing coal, coke, or oil is burned, a small amount of sulfur dioxide and sulfur trioxide will be present. For coal containing 1 to 6 wt.% sulfur, sulfur dioxide will be present in the exhaust gases in concentrations ranging from 0.05 to 0.30 vol.%.(1) When coke and coal are used, carbon monoxide (in the 1% range) and other combustible gases may be present in the exhaust gases. These emissions result mainly from uneven fuel distribution, and have been largely eliminated in the mixed feed vertical kiln. In modern oil and gas fired shaft kilns, virtually no unburnt fuel is present in the exhaust gas.(4) Nitrogen oxides will be present, regardless of the fuel burned, in concentrations ranging from 100 to 1,400 ppm.(1) The NO_x concentration will increase as the amount of excess air used in combustion increases.

Various sulfur compounds are also found in the exhaust gases from lime sludge kilns. Typical concentration ranges and averages for these compounds, which are responsible for kiln odors, are given in Table 181.(17) However, in comparison with other odor sources in the pulp mill, lime kiln odors are not significant. Hydrogen sulfide is the major odorous gas emitted from the kiln. It is caused by the presence of sodium sulfide in the lime mud and is produced by the following reaction:

$$Na_2S + CO_2 + H_2O \rightarrow Na_2CO_3 + H_2S$$

TABLE 179

TYPICAL COMPOSITIONS OF EXHAUST GASES FROM LIME SLUDGE KILNS(12)

Component	Vol.%	
	200 TPD*	292 TPD**
H_2O	37.1	30.0
CO_2	10.4	15.3
CO	0.0	0.5
O_2	3.2	0.2
N_2	49.3	54.0
	100.0	100.0

* Excess air very high, kiln exhaust temp. high (415° F).

** Excess air less than 5%, exhaust temp. 350° F.

TABLE 180

BREAKDOWN OF FUEL USAGE IN LIME ROCK KILNS DURING 1962[3]

Fuel	Percent*
Natural gas	36
Coke	30
Bituminous coal	22
Oil	6
Unspecified and miscellaneous	6
	100

*Of total 1962 fuel usage (weight) by lime rock kilns.

TABLE 181

TYPICAL GASEOUS SULFUR CONCENTRATION RANGES AND AVERAGES FOR LIME SLUDGE KILN EXHAUST(17)

Pollutant	Average (ppm)	Conc. Range (ppm)
H_2S	48	0 to 100
CH_3SH	2	0 to 3
CH_3SCH_3	1	0 to 1
$(CH_3S)_2$	0	0
SO_2	less than 10	less than 10

If the temperature is high enough, some of the hydrogen sulfide will be oxidized to sulfur dioxide, much of which is absorbed in the lime mud. This oxidation can be presented as follows:(13)

$$2H_2S + 3O_2 \rightarrow 3SO_2 + 2H_2O$$

Exit gas temperatures for chain equipped lime sludge kilns will be 300 to 400° F for long kilns and 400 to 500° F for short kilns. For short kilns without a chain system, exhaust temperatures will be as high as 1,200° F.(9)

The gas leaving lime kilns carries variable amounts of dust. The rotary lime sludge kiln exhaust will carry a greater amount of dust than the vertical lime kiln exhaust for the reasons discussed previously.

A vertical lime kiln will generally emit particles in the amount of 1% of the weight of the lime produced.(11) Actual particulate loadings have been reported to range from 0.3 to 1.0 gr/SCF(2) and, for very large kilns, from 2.5 to 4.4 gr/SCF.(4)

Particulate loadings in lime sludge kilns range from about 3 to 20 gr/SCF.(13, 14) Several compositions of the particulate matter emitted from high calcium limestone calcining kilns are shown in Tables 182 and 183,(1,3) and from lime sludge kilns in Table 184.(12,14) It is evident that the composition of the particulate matter can vary greatly. This is especially true of lime sludge kiln emissions, and is due in these cases to the variation in the composition of the lime mud feed. The most important variation in the lime mud composition is in the amount of sodium sulfide present, which is controlled by the degree of washing which the lime mud receives.

An approximate relationship between process weight and gas flow in a gas fired lime sludge kiln is given in Table 185. Table 186 shows a theoretical relationship between exhaust flow rates and vertical lime calcining kiln capacities. These rates were calculated from an actual operating example(2) and the heating content of various fuels. Excess primary combustion air was assumed to be 10%, with 30% more air required for cooling and even heat distribution. The lime produced was assumed to be 95 wt.% CaO, and the exhaust gas to contain 30.1 vol.% CO_2, 2.5 vol.% O_2 and 0 vol.% CO on a dry basis, with the remainder being nitrogen and water vapor.

AIR POLLUTION CONTROL CONSIDERATIONS

The major contaminant in vertical lime kiln exhaust gas is lime dust. This

TABLE 182

TYPICAL EMISSIONS OF A NATURAL GAS-FIRED LIME ROCK KILN(1, 3)

Component	High Calcium Wt %	Dolomitic Wt %
CaO	66.3	7.2
$CaCO_3$	23.1	64.3
$Ca(OH)_2$	6.4	–
MgO	1.4	28.2
$CaSO_4$	1.2	0.3
Other	1.6	–
	100.0	100.0

TABLE 183

TYPICAL EMISSIONS OF A PULVERIZED COAL-FIRED LIME ROCK KILN(1, 3)

Component	Wt.%
Total oxides (CaO and MgO)	37
Total carbonates ($CaCO_3$ and $MgCO_3$)	25
CaO chemically combined in compounds, including some $CaSO_4$	10
SiO_2	10
R_2O_3	5
Carbon	12
SO_3	1
Total	100

TABLE 184

VARIOUS COMPOSITIONS OF LIME SLUDGE KILN DUST

Component	Dolomitic(12) Wt.%	High Calcium(14) Wt.%
Available CaO	6.4	–
$CaCO_3$	60.8	60.3 to 91.7
Na_2CO_3	1.4	7.7 to 34.1
$MgCO_3$	18.7	–
Na_2SO_4	–	0.2 to 4.4
$Fe_2O_3 + Al_2O_3$	2.9	0.3 to 1.2
Silica and other impurities	9.8	
	100.0	100.0

TABLE 185

TYPICAL EXHAUST GAS PRODUCTION FOR SEVERAL LIME SLUDGE KILN SIZES

Lime Produced, Ton/Day	Exhaust Gas, DSCFM
60	7,150
120	14,500
200	25,000
290	31,000
400	50,000

TABLE 186

THEORETICAL LIME ROCK KILN EXHAUST RATES FOR DIFFERENT FUELS

Kiln Capacity	Exhaust Flow, SCFM (60° F)			
	Natural Gas	Fuel Oil	Bituminous Coal	Coke
25	1,100	1,280	1,400	1,500
100	4,400	5,125	5,600	6,000
250	11,000	12,800	14,000	15,000
350	15,400	17,940	19,600	21,000
700	30,800	35,875	39,200	42,000

dust is relatively large, and in many cases, the bulk of particulate matter can be removed by a multiple dry cyclone unit. Collection efficiencies range from 75 to 85%.(3) Table 187 contains a typical size distribution of particulate emissions from lime rock kilns.(11) The relatively large particle size is mainly due to the fact that there is much less attrition in a vertical kiln than in a rotary kiln.

Collection efficiencies will be higher if wet scrubbers, fabric filters, or electrostatic precipitators are used. All three types of equipment have been used successfully but precipitators have not been applied commercially in great numbers. Where high sulfur containing coal or oil is burned, wet scrubbers using an alkaline scrubbing solution have the advantage of controlling sulfur dioxide emissions as well as the particulate emissions.

The two main types of emissions from lime sludge kilns which must be controlled are the "soda fume" and the lime dust. The soda fume consists of fine particles of sodium compounds which are volatilized at the hot end of the kiln. The lime particles are larger and are entrained in the exhaust gas.

Wet scrubbing is the most widely used method for controlling these emissions. The cyclonic, variable orifice, and Venturi types of scrubbers have all been successfully employed. Diagrams of these scrubbers are shown in Figures 142, 143, and 144.(15)

Pressure drops of 15 to 17 in. w.c. across the scrubber are sufficient to remove about 99% of the lime dust but only 70 to 80% of the soda fume.(13) This small amount of soda fume emission can cause opacity problems due to the small size of the particles. The problem of sodium loss can be handled several ways. The easiest method is to properly wash the lime mud before it is burned, greatly reducing the amount of sodium sulfide put into the kiln. Electrostatic precipitators can show better soda fume collection efficiency, but, so far, they have not been widely used.

TABLE 187

TYPICAL PARTICLE SIZE DISTRIBUTION OF VERTICAL LIME ROCK KILN EMISSIONS(11)

Size Range, microns	Wt.% of Particles in Range
greater than 44	28
20 to 44	38
20 to 20	24
5 to 10	8
less than 5	2
	100

FIGURE 142

CYCLONIC GAS SCRUBBER

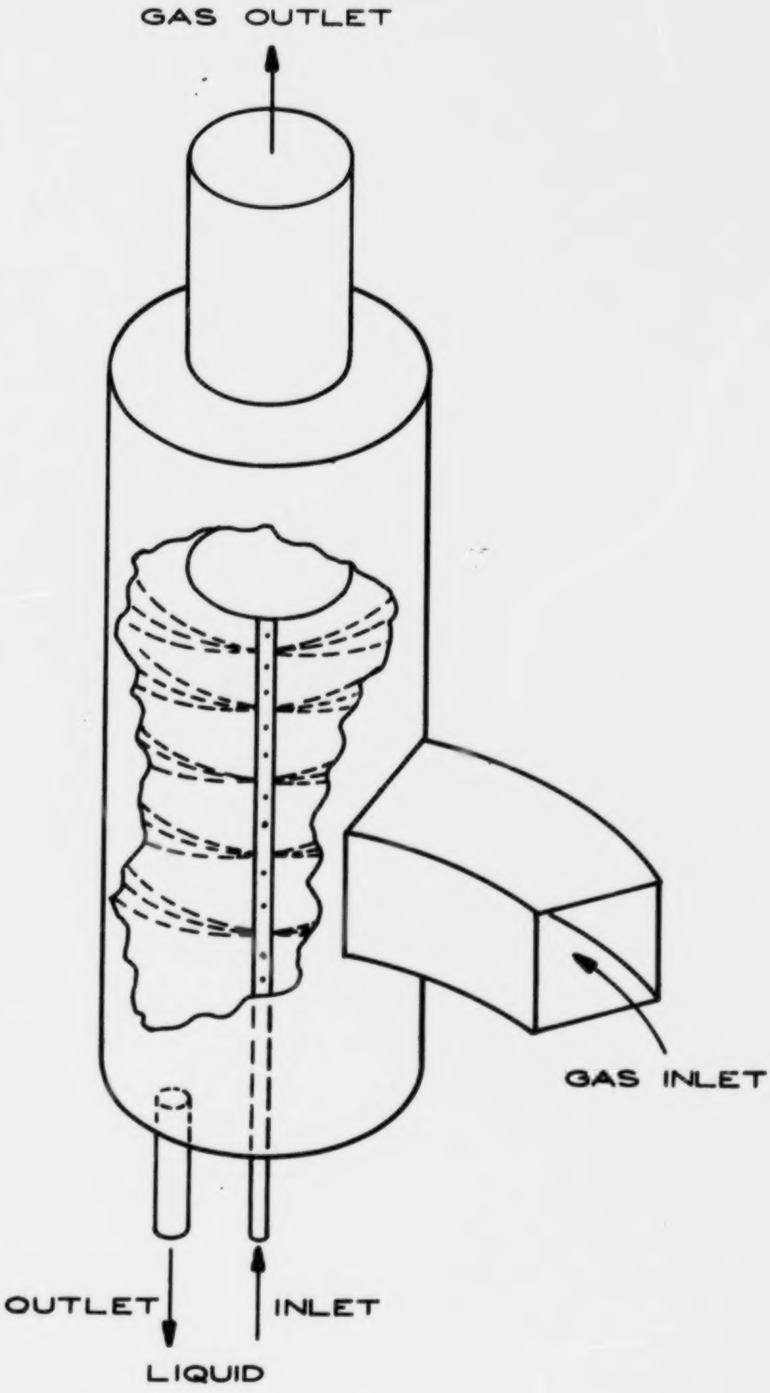

GAS OUTLET
GAS INLET
OUTLET
INLET
LIQUID

FIGURE 143

VARIABLE ORIFICE SCRUBBER

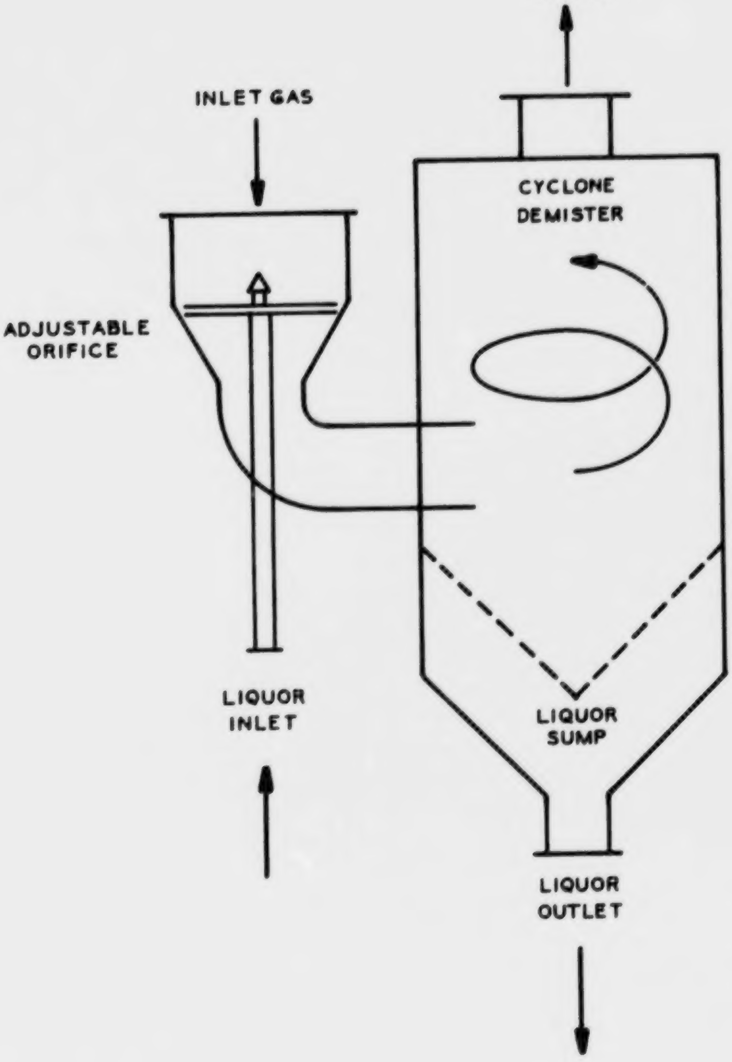
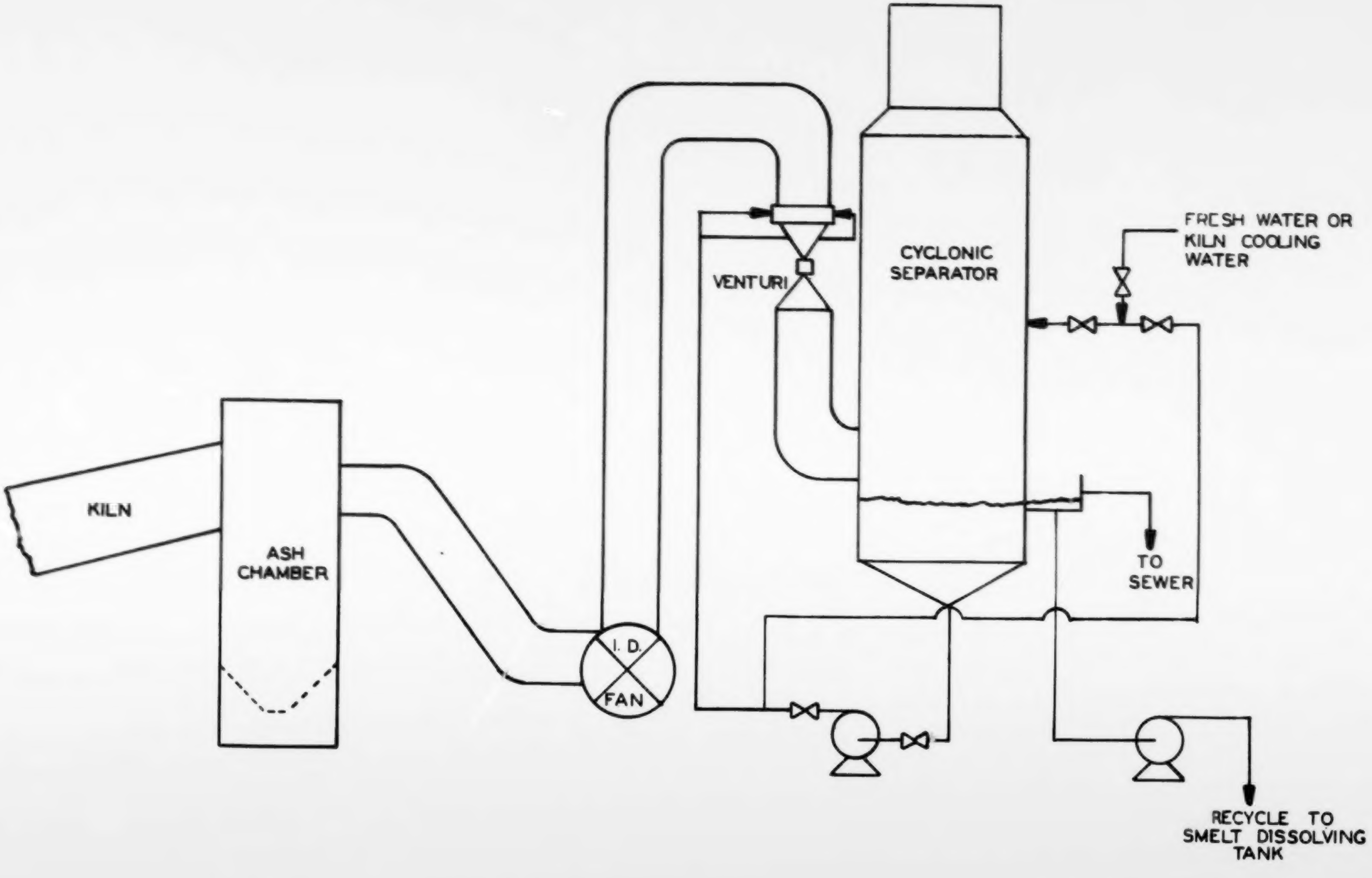

OFF GAS
INLET GAS
CYCLONE DEMISTER
ADJUSTABLE ORIFICE
LIQUOR INLET
LIQUOR SUMP
LIQUOR OUTLET

FIGURE 144

LIME KILN VENTURI SCRUBBER SYSTEM


FRESH WATER OR
KILN COOLING
WATER
CYCLONIC
SEPARATOR
VENTURI
KILN
ASH
CHAMBER
I. D.
FAN
TO
SEWER
RECYCLE TO
SMELT DISSOLVING
TANK

SPECIFICATIONS AND COSTS

Specifications have been written for three abatement systems applied to vertical lime rock kilns: wet scrubbers, fabric filters, and electrostatic precipitators. The model plant sizes chosen for the precipitator specifications were larger than those chosen for both the fabric filter and wet scrubber systems. This was done because precipitators are economically attractive for only the largest sized vertical kilns. Cost data and equipment specifications for vertical kilns are presented in Tables 188 to 201 and Figures 145 to 156.

Two systems were specified for rotary sludge kilns in paper plants: wet scrubbers and electrostatic precipitators. Since gas flows for rotary sludge kilns are higher than those for vertical kilns, the model plant sizes chosen for both abatement systems were the same. A fabric filter system was not applied because the high humidity of the exhaust gas from the kiln causes blinding problems. Cost data and equipment specifications are presented in Tables 202 to 211 and Figures 157 to 166.

TABLE 188

WET SCRUBBER PROCESS DESCRIPTION FOR VERTICAL LIME ROCK KILN SPECIFICATION

The scrubber is to remove entrained limestone and lime dust from the exhaust gas of a vertical lime rock kiln. The kiln is fired with natural gas. A portion of the hot exhaust gas from the calcining zone is recirculated for heat recovery. The kiln is fed with 6 to 8 in. sized pieces of high calcium limestone.

The exhaust gas is to be brought from the kiln exhaust ports to a fan located 20 feet outside of the kiln enclosure. The fan outlet is to be five feet above grade. The scrubber will be located beyond the fan in an area free of space limitations. The scrubber should be designed to withstand the discharge pressure developed by the fan. Fresh makeup water is available and is to be added to the recirculation tank. The scrubber is to operate so as to reduce continuously the kiln outlet loading to the levels specified.

The scrubbing system should include the following:

1. *Venturi scrubber with a cyclonic entrainment separator.*
2. *Recirculation tank and pumps.*
3. *Slurry settler, which will handle a portion of the recirculation pump discharge, capable of producing a reasonably thickened underflow product while returning water fully treated to minimize solids content. Slurry withdrawal should be set to maintain 10% (by weight) solids when the kiln is operating at design capacity.*
4. *Minimum of two filters to dewater the slurry product, capable of producing a cake with a minimum of 65% (by weight) solids.*
5. *Necessary fans, dampers and motors.*
6. *Necessary controls.*
7. *Carbon steel construction.*
8. *Packing glands flushed with fresh water to prevent binding of the seals.*

TABLE 189

WET SCRUBBER OPERATING CONDITIONS FOR VERTICAL LIME ROCK KILN SPECIFICATION

Two sizes of wet scrubbers are to be quoted for each of two efficiency levels. Vendors' quotations should consist of four separate and independent quotations.

	Small	Large
Kiln Capacity, ton/day	100	350
Process wt., lb/hr	15,000	51,000
Inlet Gas		
Flow, ACFM	7,300	25,500
Temp., °F	275	275
Flow, SCFM	5,300	18,400
% Moisture (vol)	12	12
Inlet Solids Rate, lb/hr	80.7	288.5
Inlet Solids Loading		
gr/ACF	1.29	1.32
gr/DSCF	2.03	2.08
Outlet Gas		
Flow, ACFM	6,500	22,700
Temp., °F	130	130
% Moisture (vol)	15	15
Case 1 – Medium Efficiency*		
Outlet Solids Rate, lb/hr	15.8	35.9
Outlet Loading		
gr/ACF	0.28	0.18
gr/DSCF	0.37	0.24
Collection Efficiency, wt.%	80.4	87.6
Scrubber Δ P	12 in. w.c.	12 in. w.c.
Case 2 – High Efficiency*		
Outlet Solids Rate, lb/hr	0.56	1.94
Outlet Solids Loading, gr/ACF	0.01	0.01
Outlet Solids Loading, gr/DSCF	<0.015	<0.015
Collection Efficiency, wt.%	99.31	99.33
Scrubber Δ P	30 in. w.c.	30 in. w.c.

*See page 6 for definition of collection efficiency levels.

TABLE 190

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS

	LA Process Wt.		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	7,300	25,500	7,300	25,500
°F	275	275	275	275
SCFM	5,300	18,400	5,300	18,400
Moisture Content, Vol. %	12	12	12	12
Effluent Contaminant Loading				
gr/ACF	1.29	1.32	1.29	1.32
lb/hr	80.7	288.5	80.7	288.5
Cleaned Gas Flow				
ACFM	6,500	22,700	6,500	22,700
°F	130	130	130	130
SCFM	5,850	20,400	5,850	20,400
Moisture Content, Vol. %	15	15	15	15
Cleaned Gas Contaminant Loading				
gr/ACF	0.28	0.18	0.01	0.01
lb/hr	15.8	35.9	0.56	1.94
Cleaning Efficiency, %	80.4	87.6	99.31	99.33
(1) Gas Cleaning Device Cost	6,699	13,141	6,813	13,357
(2) Auxiliaries Cost	4,407	9,426	6,559	11,546
(a) Fan(s)				
(b) Pump(s)				
(c) Damper(s)				
* (d) Conditioning, Equipment				
* (e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering	22,767	35,717	25,100	38,083
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost	33,873	58,284	38,472	62,986

TABLE 191

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year		8,000	8,000	8,000	8,000
Operating Labor (if any)					
Operator					
Supervisor					
Total Operating Labor					
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		1,123	1,840	1,290	2,007
Replacement Parts		619	1,179	727	1,283
Total Replacement Parts		619	1,179	727	1,283
Utilities					
Electric Power	$.011/kw-hr	3,448	10,912	4,841	15,726
Fuel					
Water (Process)	$.25/M gal	191	647	191	647
Water (Cooling)					
Chemicals, Specify					
Total Utilities		3,639	11,559	5,032	16,373
Total Direct Cost		5,381	14,578	7,049	19,663
Annualized Capital Charges		3,387	5,828	3,847	6,299
Total Annual Cost		8,768	20,406	10,896	25,962

FIGURE 145

CAPITAL COSTS FOR WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS (MEDIUM EFFICIENCY)

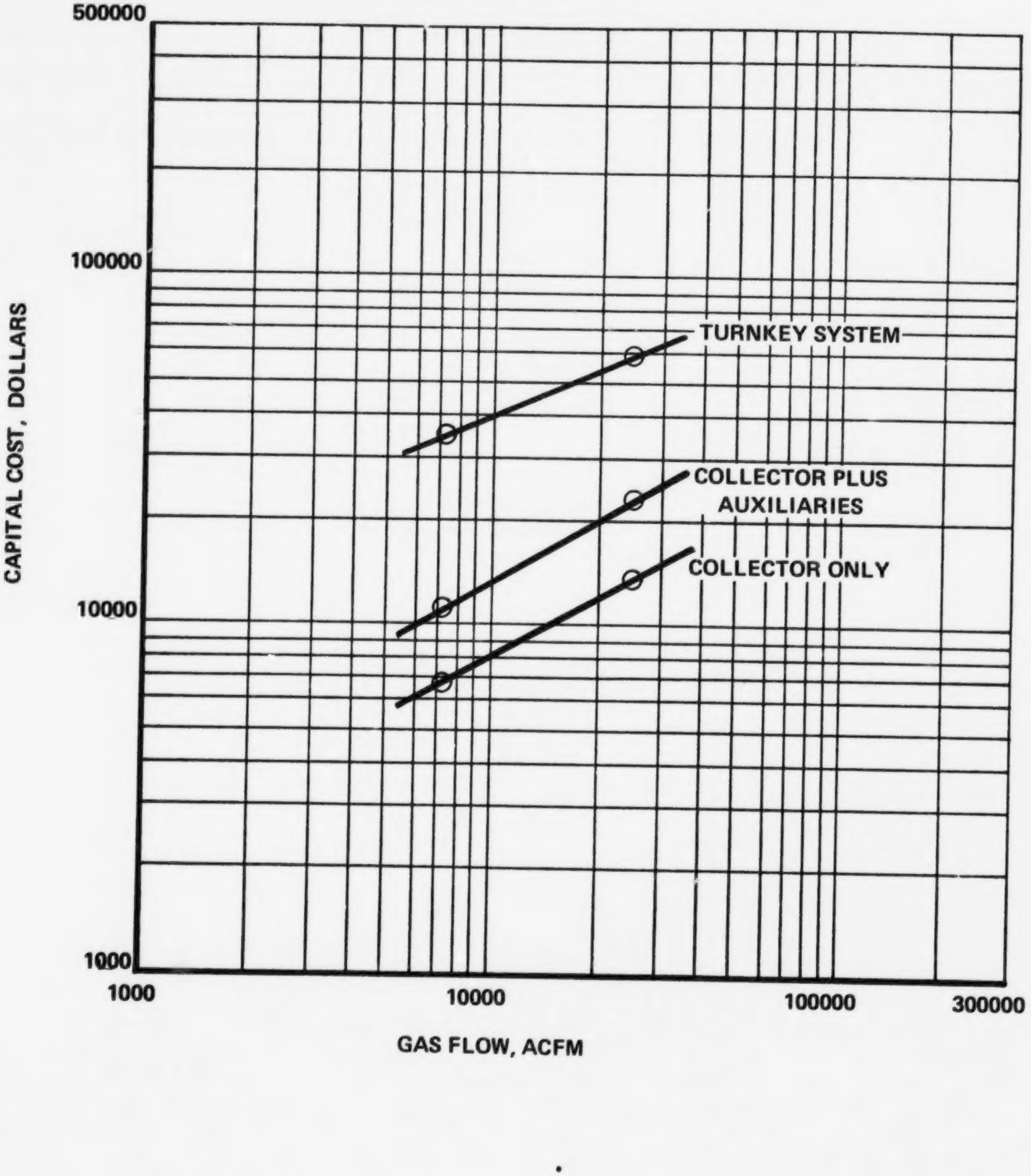


FIGURE 146

ANNUAL COSTS FOR WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS (MEDIUM EFFICIENCY)

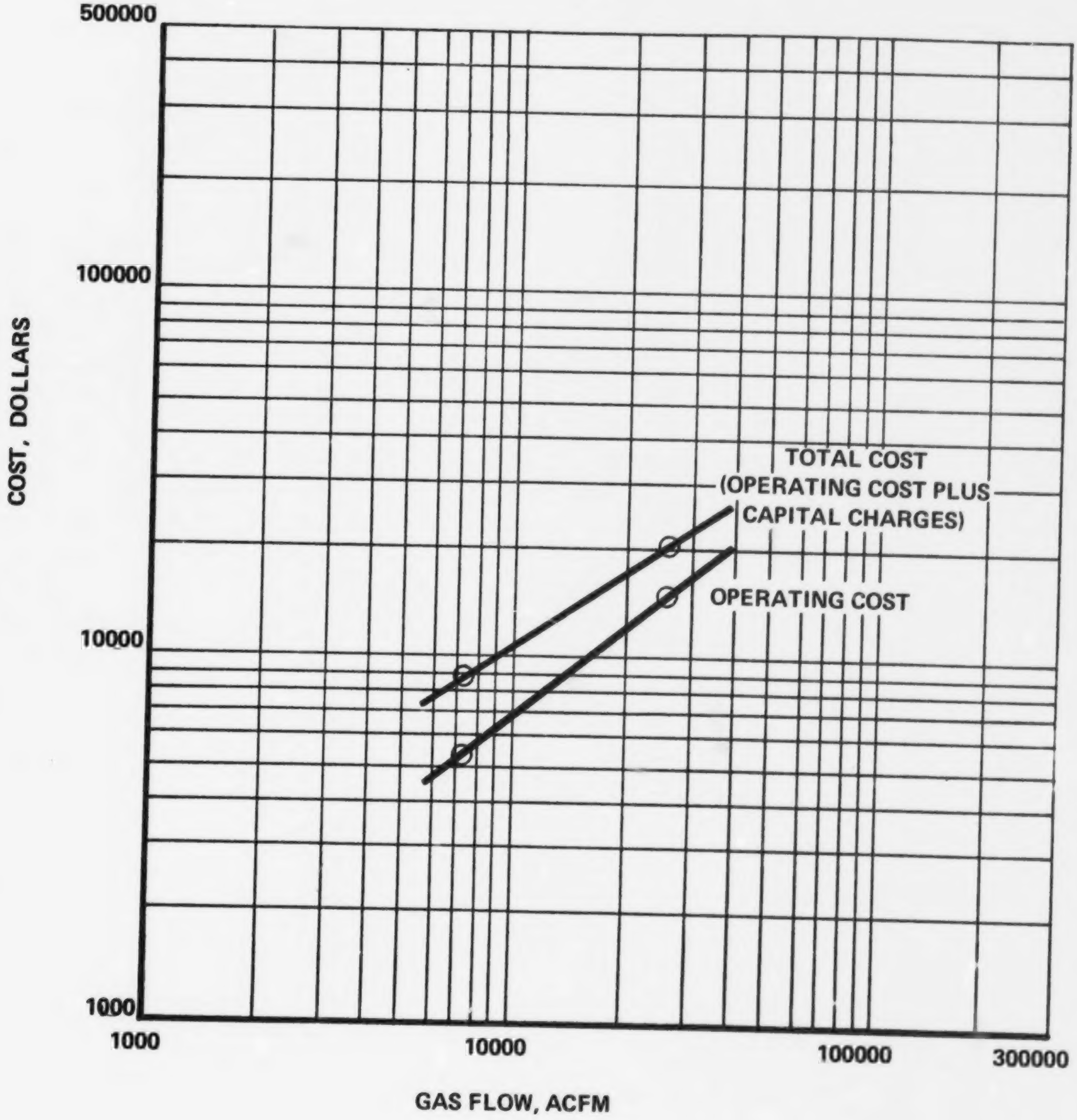


FIGURE 147

CAPITAL COSTS FOR WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

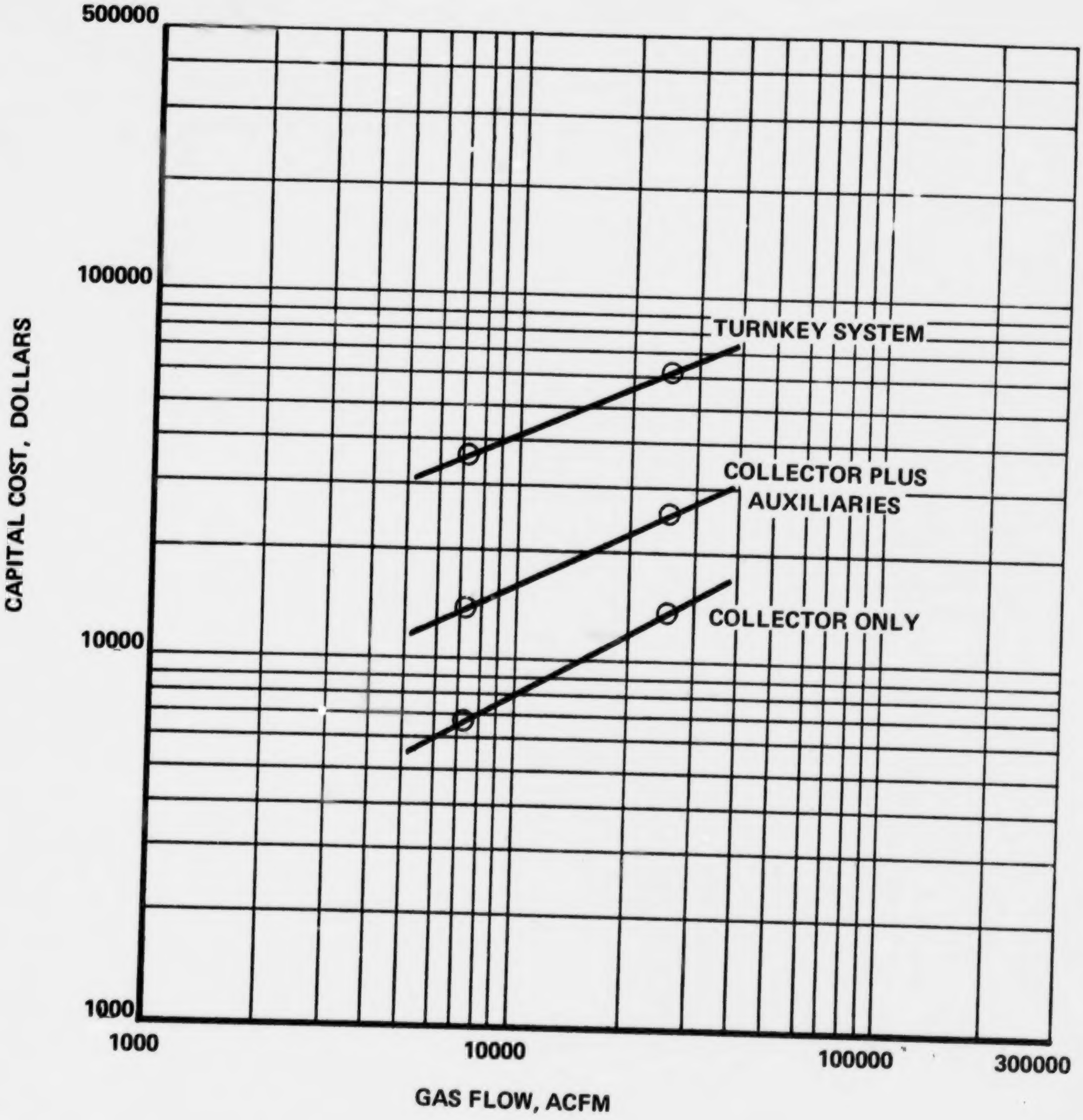


FIGURE 148

ANNUAL COSTS FOR WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

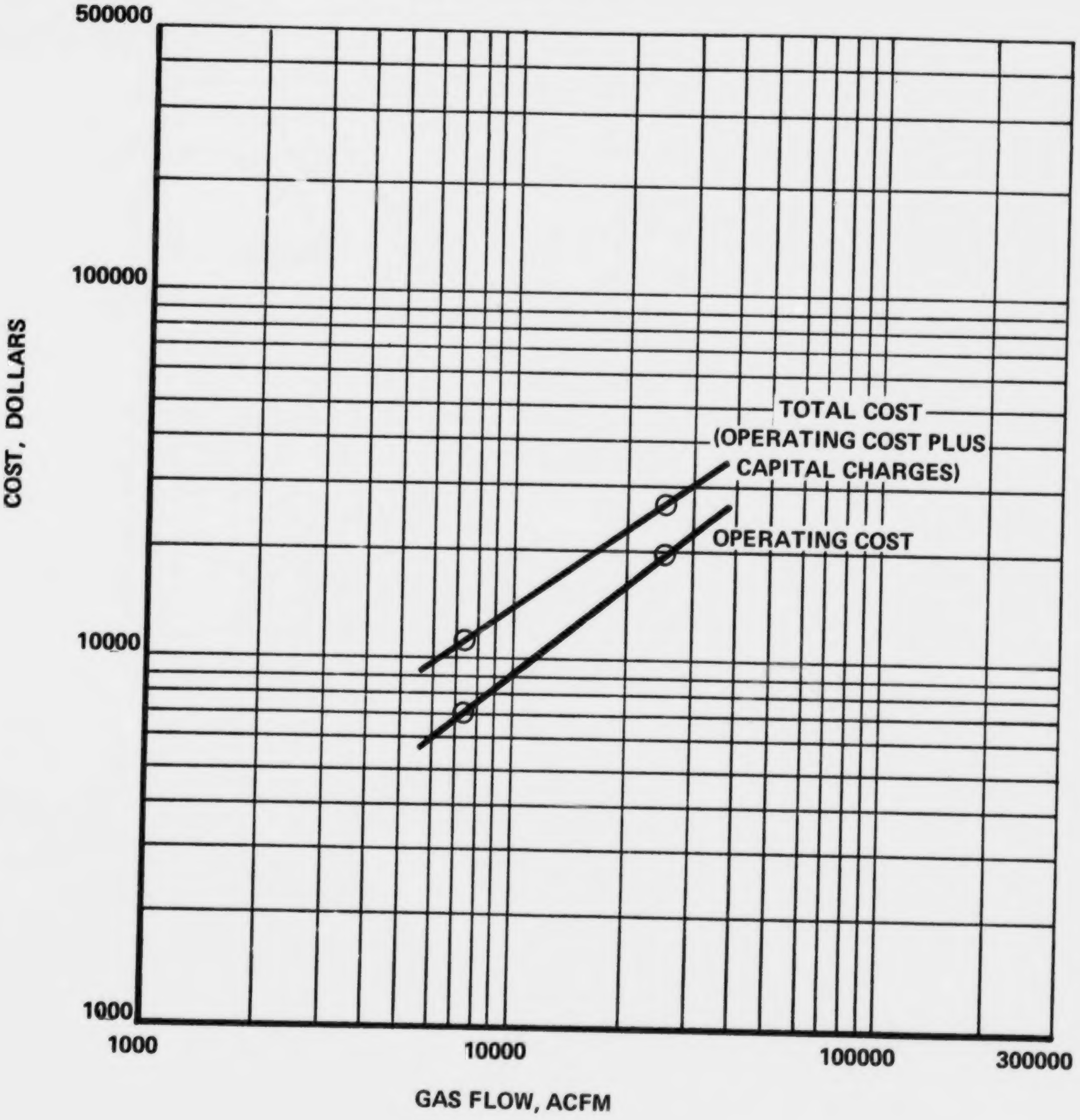


TABLE 192

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

Population Size – 20　　Sample Size – 3

Capital Cost = $38,471

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$ 31,783	$ 45,160
75	25,512	51,431
90	15,862	61,080
95	6,940	70,003

Capital Cost = $62,986

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$ 50,488	$ 75,485
75	38,770	87,203
90	20,739	105,234
95	4,065	121,908

FIGURE 149

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

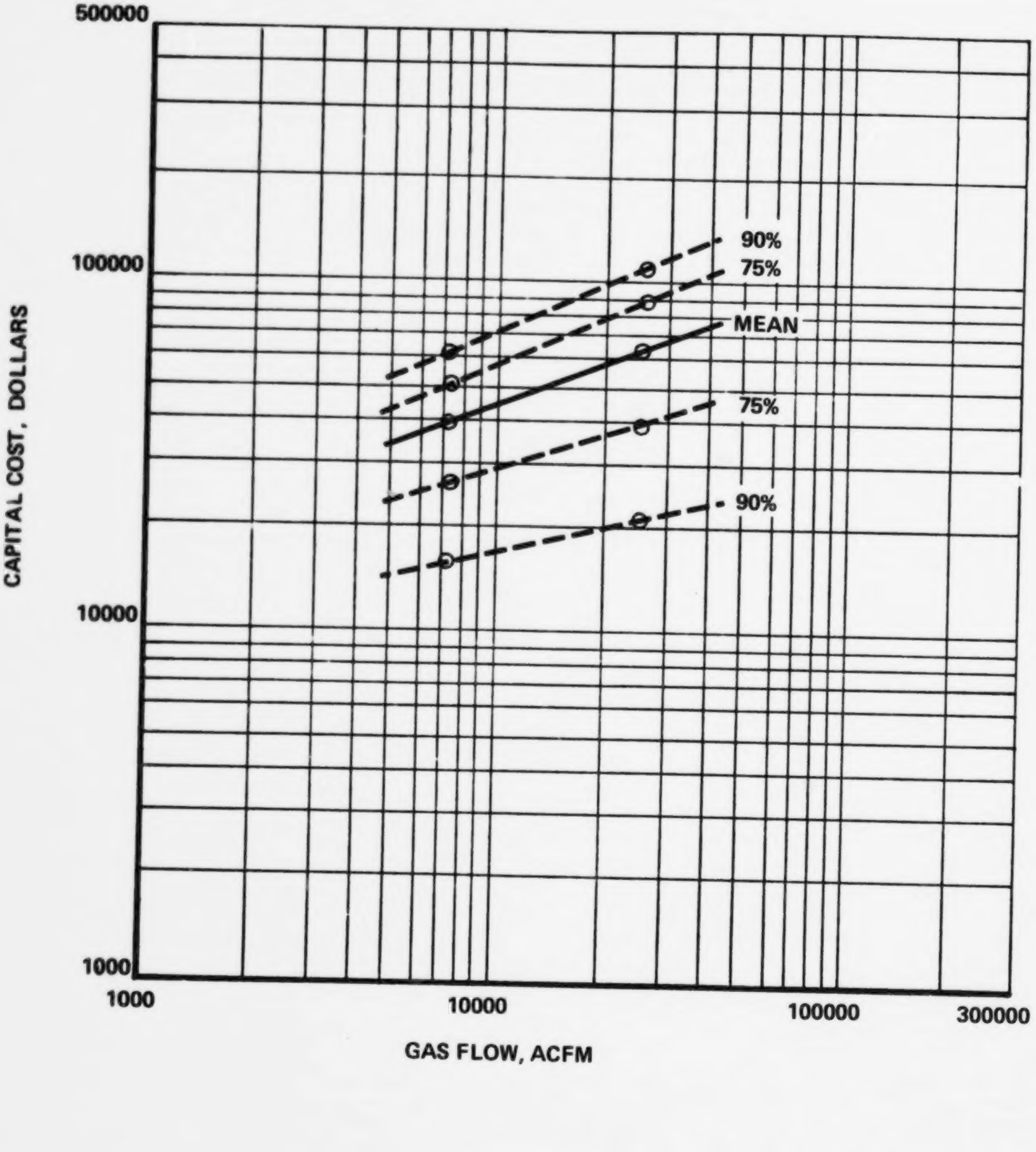


TABLE 193

FABRIC FILTER PROCESS DESCRIPTION FOR VERTICAL LIME ROCK KILN SPECIFICATION

The fabric filter is to remove entrained limestone and lime dust from the exhaust gas of a vertical lime rock kiln. The kiln is fired with natural gas. A portion of the hot exhaust gas from the calcining zone is recirculated for heat recovery. The kiln is fed with 6 to 8 in. sized pieces of high calcium limestone.

The exhaust gas is to be brought from the kiln exhaust ports to a location 20 feet outside of the kiln enclosure by means of a fan. The fabric filter will be located in an area free of space limitations. The fan is to follow the filter and the fan outlet is to be five feet above grade.

The fabric filter is to be compartmented to allow for isolation of an individual compartment for cleaning during operation. A single compartment should have a maximum of 25% of the total collecting surface area. Each section should also be capable of isolation for maintenance and have provisions for personal safety during filter operation. No more than two bags must be removed to permit access to all of the bags. The dust collecting process should be continuous and should include the following:

1. *Compartmented fabric filter operating with negative pressure.*
2. *Maximum air to cloth ratio, when one compartment is down for cleaning, of 1.80.*
3. *Fiberglass bags.*
4. *Reverse air type cleaning system. This air is to be recirculated.*
5. *Insulation of entire filter system.*
6. *Trough hoppers with a minimum side and valley angle of 67.5°. The hoppers should be capable of retaining the dust collected over 24 hours of normal operation.*
7. *Oversized screw conveyor system which includes a 9 in. diameter (min) conveyor with a rotary air lock at its end.*
8. *Dust bin sized for seven days storage capacity which is to be adjacent to the filter and have a 15 foot clearance from grade.*
9. *Necessary fans and motors.*

TABLE 194

FABRIC FILTER OPERATING CONDITIONS FOR

VERTICAL LIME ROCK KILN SPECIFICATION

Two sizes of fabric filters are specified for the "high efficiency" level. Vendors' quotations should consist of one quotation for each of the two sizes, with a representation of the efficiency expected for the unit quoted. The efficiency quoted may be better than the "high efficiency" case.

	Small	*Large*
Kiln Capacity, ton/day	100	350
Process wt., lb/hr	15,000	51,000
Inlet Gas		
Flow, ACFM	7,300	25,500
Temp., °F	275	275
Flow, SCFM	5,300	18,400
% Moisture (vol)	12	12
Dew Point, °F	122	122
Inlet Solids Rate, lb/hr	80.7	288.5
Inlet Solids Loading, gr/ACF	1.29	1.32
Inlet Solids Loading, gr/DSCF	2.02	2.02
High Efficiency		
Outlet Solids Rate, lb/hr	0.63	2.19
Outlet Solids Loading, gr/ACF	0.01	0.01
Outlet Solids Loading, gr/DSCF	0.015	0.015
Collector Efficiency, wt.%	99.22	99.24
Air-To-Cloth Ratio	1.5/1	1.5/1

TABLE 195

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR FABRIC FILTERS FOR VERTICAL LIME ROCK KILNS

	LA Process Wt.		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM			7,300	25,500
°F			275	275
SCFM			5,300	18,400
Moisture Content, Vol. %			12	12
Effluent Contaminant Loading				
gr/ACF			1.29	1.32
lb/hr			80.7	288.5
Cleaned Gas Flow				
ACFM			7,300	25,500
°F			275	275
SCFM			5,300	18,400
Moisture Content, Vol. %			12	12
Cleaned Gas Contaminant Loading				
gr/ACF			0.01	0.01
lb/hr			0.63	2.19
Cleaning Efficiency, %			99.22	99.24
(1) Gas Cleaning Device Cost			30,573	78,427
(2) Auxiliaries Cost				
(a) Fan(s)			2,556	6,213
(b) Pump(s)				
(c) Damper(s) Emer.Temp.air			2,000	2,000
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment			5,850	5,850
(3) Installation Cost				
(a) Engineering			29,234	52,318
(b) Foundations & Support				
Ductwork				
Stack				
*Electrical				
Piping				
*Insulation				
Painting				
Supervision				
*Startup				
Performance Test				
Other				
(4) Total Cost			70,213	144,808

*Not included.

FIGURE 150

CAPITAL COSTS FOR FABRIC FILTERS FOR VERTICAL LIME ROCK KILNS

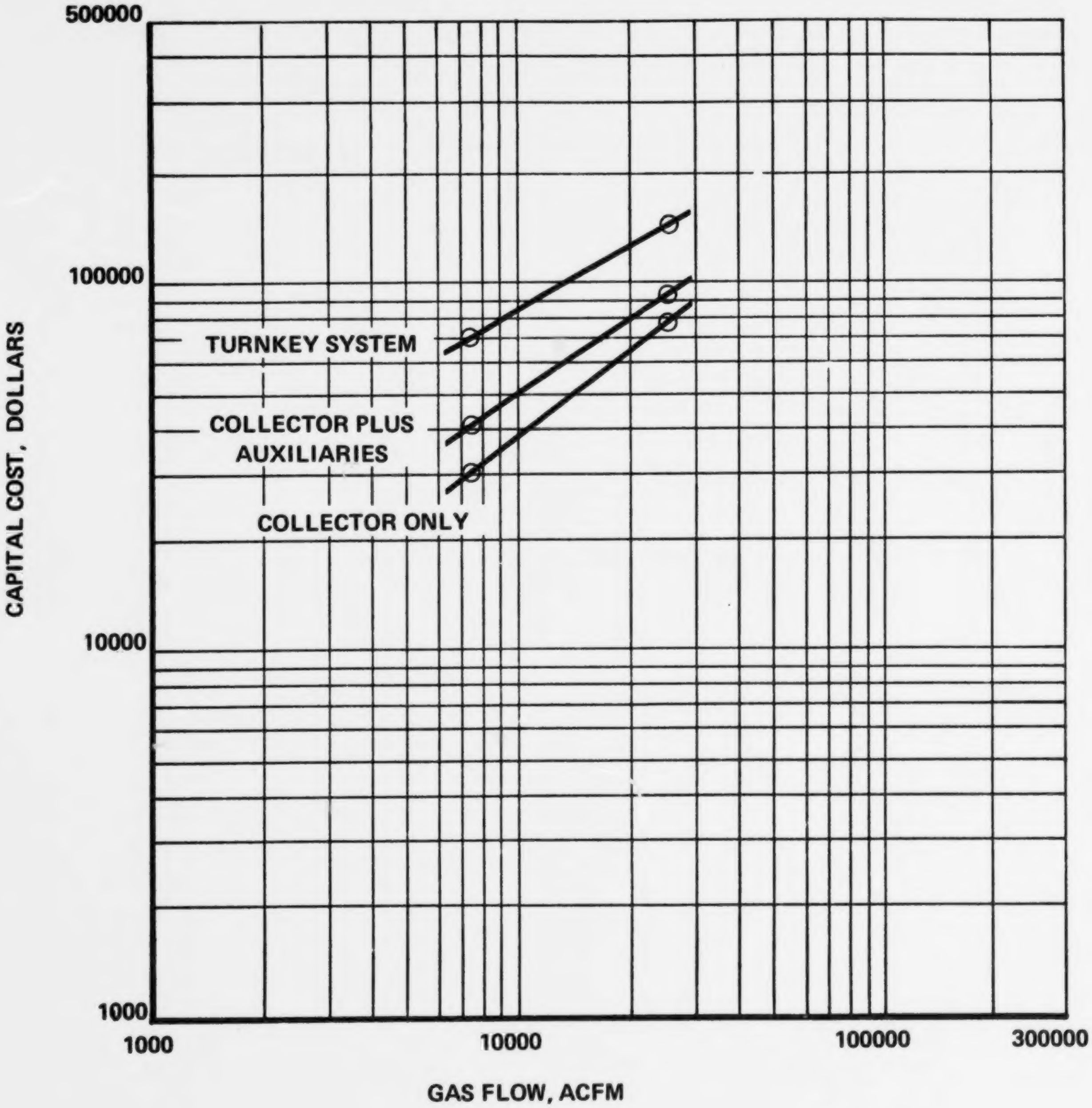


TABLE 196

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR FABRIC FILTERS FOR VERTICAL LIME ROCK KILNS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year				8,000	8,000
Operating Labor (if any)					
Operator					
Supervisor					
Total Operating Labor					
Maintenance					
Labor	$6/hr			9,000	12,000
Materials				200	600
Total Maintenance				9,200	12,600
Replacement Parts				400	1,400
Total Replacement Parts				400	1,400
Utilities					
Electric Power	$.011/kw-hr			1,980	6,600
Fuel					
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities				1,980	6,600
Total Direct Cost				11,580	20,600
Annualized Capital Charges				6,821	14,481
Total Annual Cost				18,401	35,081

FIGURE 151

ANNUAL COSTS FOR FABRIC FILTERS FOR VERTICAL LIME ROCK KILNS

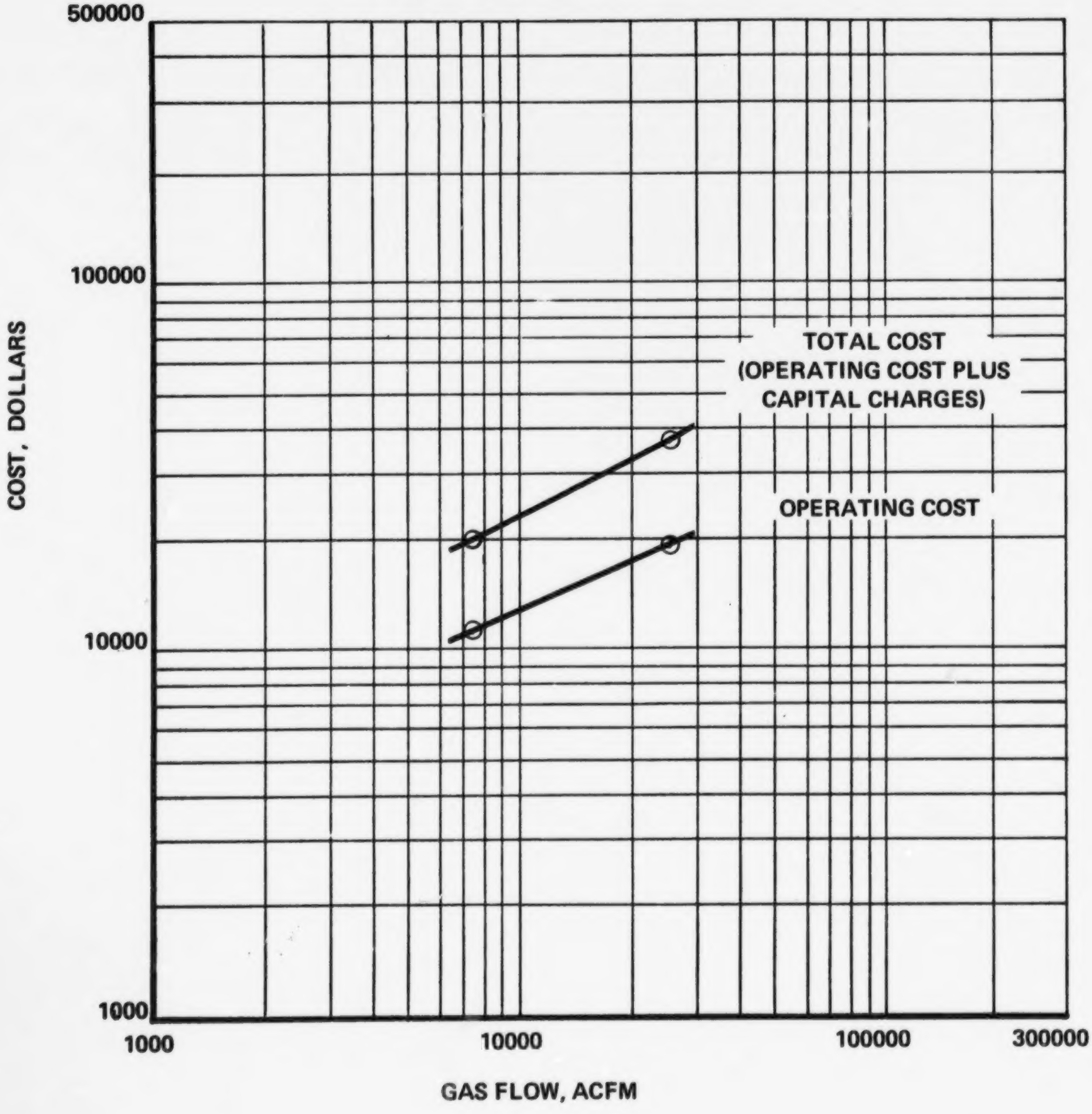


TABLE 197

ELECTROSTATIC PRECIPITATOR PROCESS DESCRIPTION FOR VERTICAL LIME ROCK KILN SPECIFICATION

A single electrostatic precipitator is to remove entrained limestone and lime dust from the exhaust gas of a vertical lime rock kiln. The kiln is fired with natural gas. A portion of the hot exhaust gas from the calcining zone is recirculated for heat recovery. The kiln is fed with 6 to 8 in. sized pieces of high calcium limestone.

The exhaust gas is to be brought from the kiln exhaust ports to a location 20 feet outside of the kiln enclosure and 20 feet above grade by means of a fan. The precipitator will be at ground level in an area beyond the ductwork which is free of space limitations. The fan will follow the precipitator. The precipitator is to reduce continuously the kiln outlet loading to the levels specified.

The precipitator system should include the following:

1. *Precipitator provided with a minimum of two independent electrical fields in the direction of gas flow.*
2. *Trough type hoppers equipped with continuous dust removal by screw conveyor to a dust tank. The conveying system must be provided with suitable sealing for negative pressure operation.*
3. *Automatic voltage control.*
4. *Safety interlocked system which prevents access to the interior of the precipitator unless the electrical circuitry is disconnected and grounded.*
5. *Rapping system which is adjustable in terms of both intensity and rapping period.*
6. *Necessary fans and motors.*
7. *Model study for precipitator gas distribution.*

TABLE 198

ELECTROSTATIC PRECIPITATOR OPERATING CONDITIONS FOR VERTICAL LIME ROCK KILN SPECIFICATION

Two sizes of precipitators are to be quoted for each of two efficiency levels. Vendors' quotations should consist of four separate and independent quotations.

	Small	*Large*
Kiln Capacity, ton/day	*350*	*700*
Process wt., lb/hr	*51,000*	*102,000*
Inlet Gas		
Flow, ACFM	*25,500*	*50,900*
Temp., °F	*275*	*275*
Flow, SCFM	*18,400*	*36,700*
% Moisture (vol)	*12*	*12*
Dew Point, °F	*122*	*122*
Inlet Solids Rate, lb/hr	*288.5*	*615.2*
Inlet Solids Loading, gr/ACF	*1.32*	*1.41*
Inlet Solids Loading, gr/DSCF	*2.08*	*2.08*
*Case 1 – Medium Efficiency**		
Outlet Solids Rate, lb/hr	*35.9*	*40.0*
Outlet Solids Loading, gr/ACF	*0.16*	*0.09*
Outlet Solids Loading, gr/DSCF	*0.25*	*0.14*
Collection Efficiency, wt.%	*87.6*	*93.5*
Drift Velocity, fps	*0.25*	*0.25*
*Case 2 – High Efficiency**		
Outlet Solids Rate, lb/hr	*2.19*	*4.36*
Outlet Solids Loading, gr/ACF	*0.01*	*0.01*
Outlet Solids Loading, gr/DSCF	*0.015*	*0.015*
Collection Efficiency, wt.%	*99.24*	*99.29*
Drift Velocity, fps	*0.25*	*0.25*

**See page 6 for definition of collection efficiency levels.*

TABLE 199

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS

	LA Process Wt.		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	25,500	50,900	25,500	50,900
°F	275	275	275	275
SCFM	18,400	36,700	18,400	36,700
Moisture Content, Vol. %	12	12	12	12
Effluent Contaminant Loading				
gr/ACF	1.32	1.41	1.32	1.41
lb/hr	288.5	615	288.5	615
Cleaned Gas Flow				
ACFM	25,500	50,900	25,500	50,900
°F	275	275	275	275
SCFM	18,400	36,700	18,400	36,700
Moisture Content, Vol. %	12	12	12	12
Cleaned Gas Contaminant Loading				
gr/ACF	0.16	0.09	0.01	0.01
lb/hr	35.9	40	2.19	4.36
Cleaning Efficiency, %	87.6	93.5	99.24	99.29
(1) Gas Cleaning Device Cost	52,705	71,995	67,745	95,515
(2) Auxiliaries Cost	23,713	36,665	25,838	38,225
(a) Fan(s)				
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering	45,140	68,400	60,060	82,350
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost	121,558	177,060	153,643	216,090

TABLE 200

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)					
Operator		2,000	2,000	2,000	2,000
Supervisor	$6/hr				
Total Operating Labor		2,000	2,000	2,000	2,000
Maintenance					
Labor		1,072	2,139	1,072	2,139
Materials	$6/hr				
Total Maintenance		1,072	2,139	1,072	2,139
Replacement Parts		600	1,200	600	1,200
Total Replacement Parts		600	1,200	600	1,200
Utilities					
Electric Power	$.011/kw-hr	1,320	3,080	3,036	5,984
Fuel					
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities		1,320	3,080	3,036	5,984
Total Direct Cost		4,992	8,419	6,708	11,323
Annualized Capital Charges		12,156	17,706	15,364	21,609
Total Annual Cost		17,148	26,125	22,072	32,932

FIGURE 152

CAPITAL COSTS FOR ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS (MEDIUM EFFICIENCY)

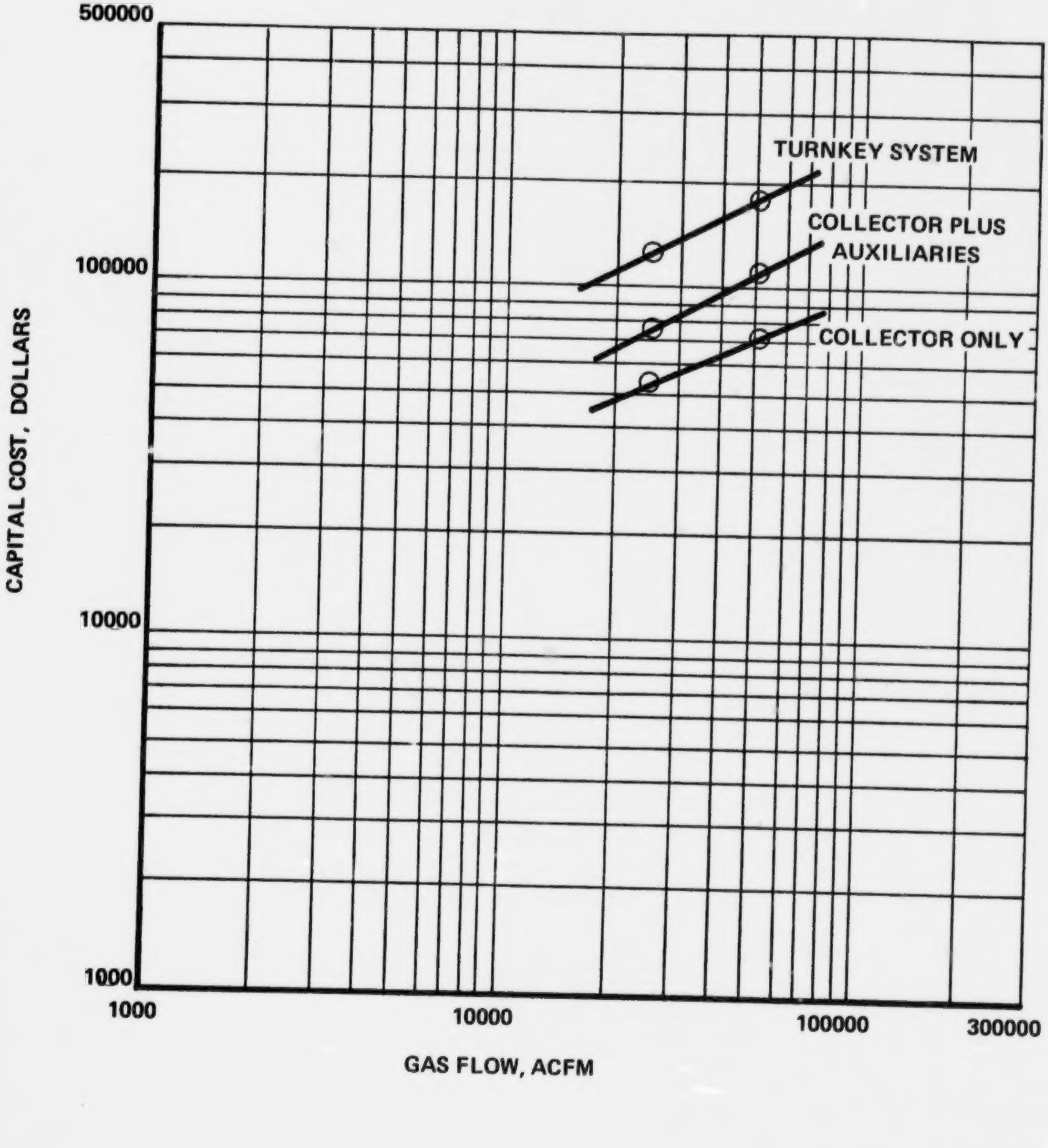


FIGURE 153

ANNUAL COSTS FOR ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS (MEDIUM EFFICIENCY)

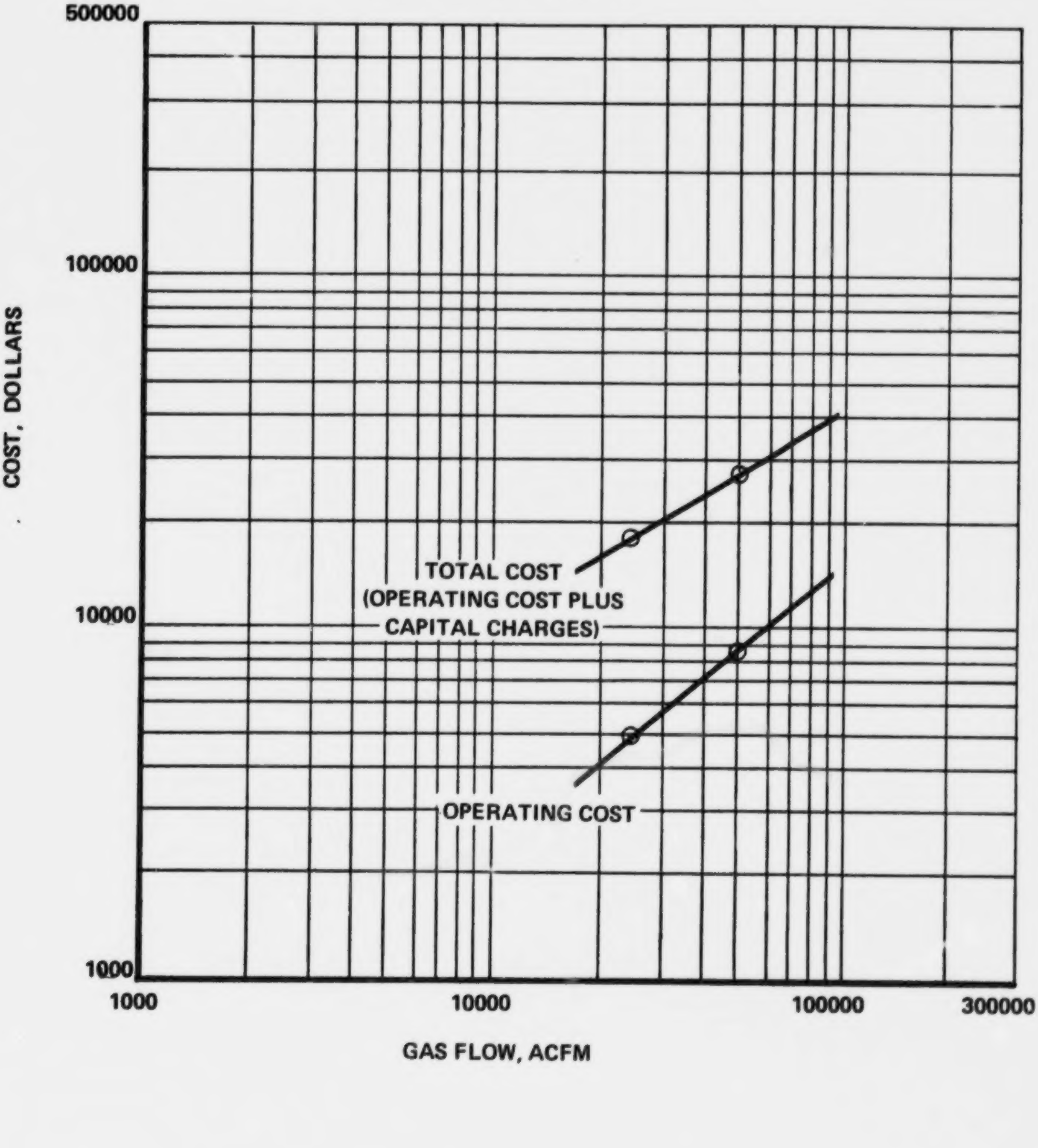


FIGURE 154

CAPITAL COSTS FOR ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

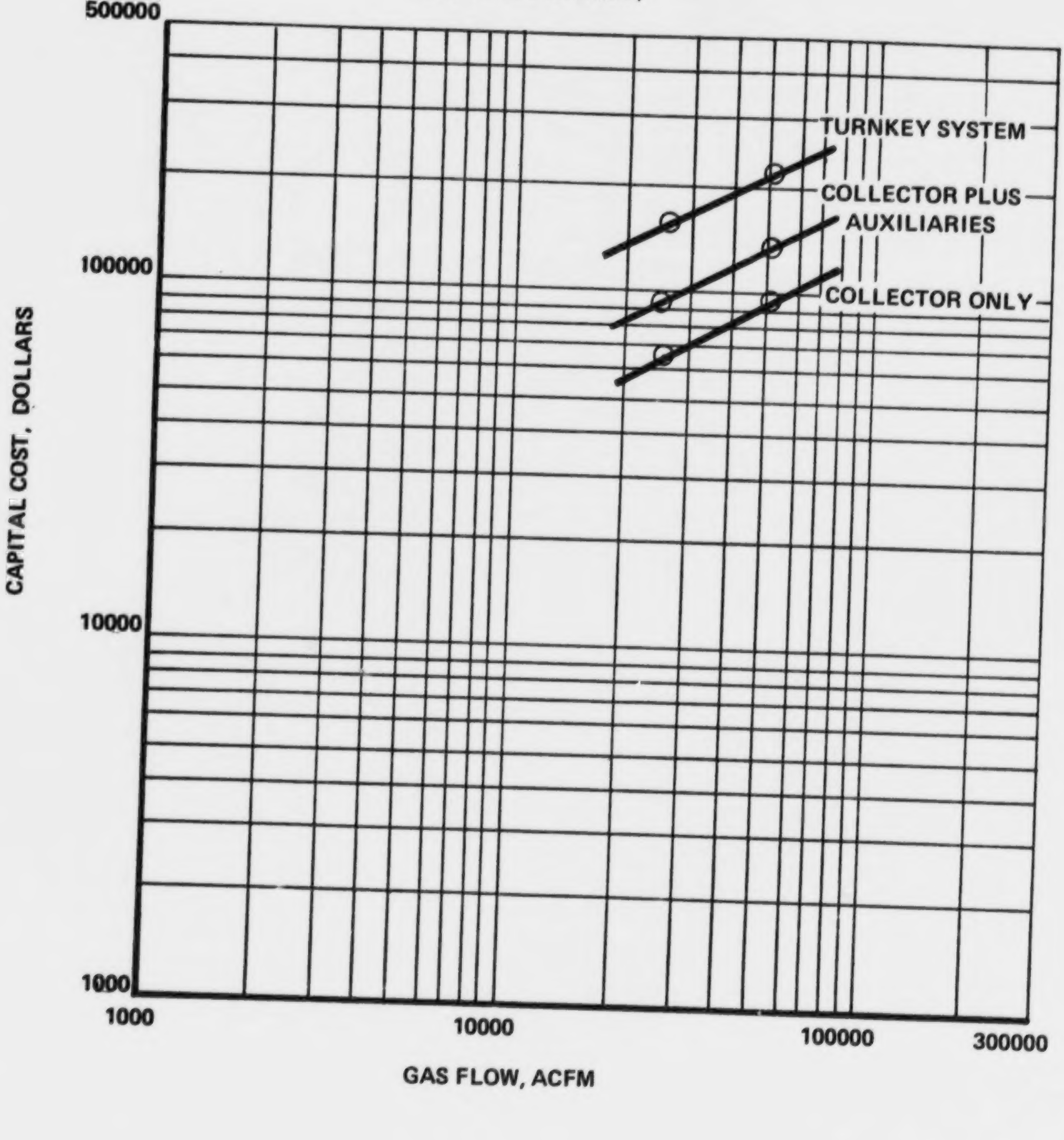


FIGURE 155

ANNUAL COSTS FOR ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

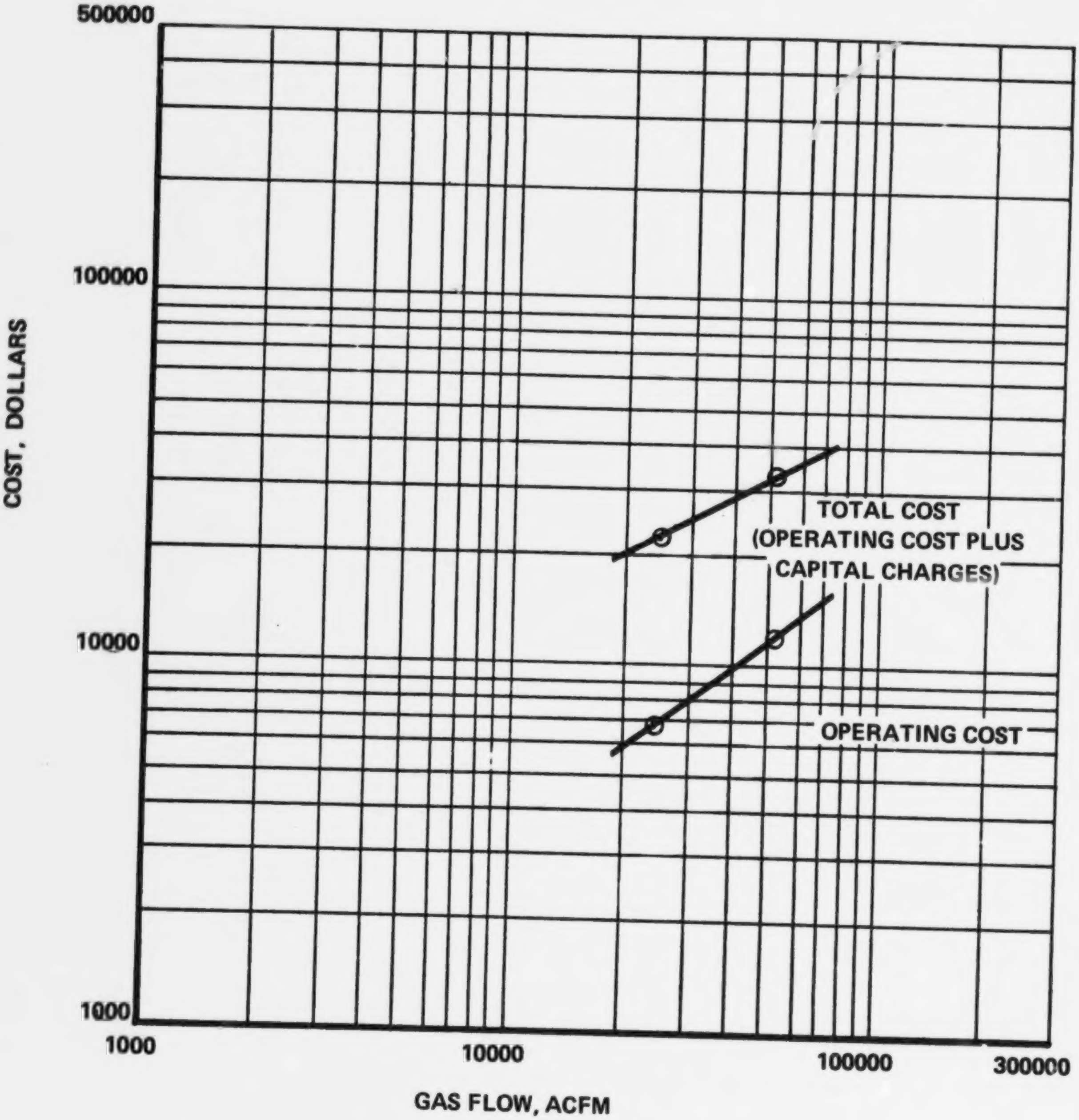


TABLE 201

CONFIDENCE LIMITS FOR CAPITAL COST OF ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

Population Size – 20 Sample Size – 2

Capital Cost = $153,643

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$143,515	$163,770
75	131,215	176,070
90	107,138	200,147
95	81,134	226,151

Capital Cost - $216,090

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$185,127	$247,053
75	147,521	284,659
90	73,910	358,270
95	5,594	437,774

FIGURE 156

CONFIDENCE LIMITS FOR CAPITAL COST OF ELECTROSTATIC PRECIPITATORS FOR VERTICAL LIME ROCK KILNS (HIGH EFFICIENCY)

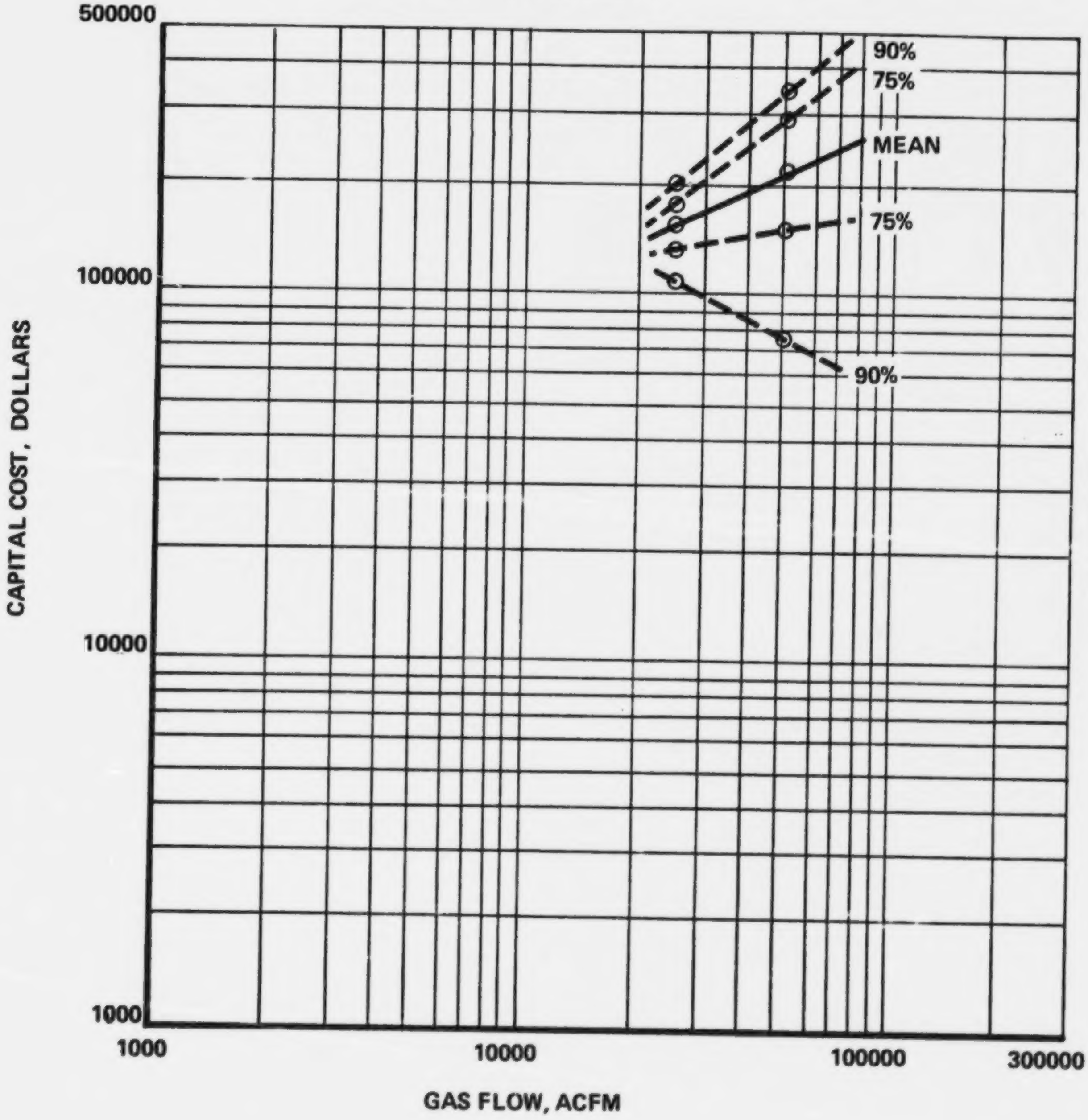


TABLE 202

WET SCRUBBER PROCESS DESCRIPTION FOR ROTARY LIME SLUDGE KILN SPECIFICATION

The scrubber is to remove entrained lime dust and soda fume from the exhaust gas of a rotary kiln used to produce reburned lime for the sulfate pulping process. The kiln is fired with natural gas. The kiln is fed with lime sludge containing 55 wt.% solids. If the sludge has been washed properly, it will contain about 0.5 wt.% sodium, expressed as Na_2O on a dry lime mud basis. However, this is not always the case and the sludge to this kiln has a sodium content, expressed as Na_2O, of about 2.0 wt.%. The feed end of the kiln is equipped with a dust fallout chamber and a heat recuperating chain system.

The exhaust gas will be brought from the feed end housing to a fan located 20 feet to one side of the kiln enclosure. The fan outlet is to be five feet above grade. The scrubber will be located beyond the fan in an area free of space limitations. The scrubber should be designed to withstand the discharge pressure developed by the fan. Clean makeup water is to be added to the recirculation tank. The scrubber is to operate so as to reduce continuously the kiln outlet loading to the levels specified.

The scrubbing system should include the following:

1. *Scrubber with a variable throat Venturi, flooded elbow, and a cyclonic entrainment separator.*
2. *Fifty foot stack following the scrubber.*
3. *Recirculation tank and pumps.*
4. *Slurry settler, which will handle a portion of the recirculation pump discharge, capable of producing a reasonably thickened underflow product while returning water fully treated to minimize solids content. Slurry withdrawal should be set to maintain 10% (by weight) solids when the kiln is operating at design capacity.*
5. *Minimum of two filters to dewater the slurry product, capable of producing a cake with a minimum of 65% (by weight) solids.*
6. *Necessary fans, dampers, and motors.*
7. *Necessary controls.*
8. *Carbon steel construction.*
9. *Packing glands flushed with fresh water to prevent binding of the seals.*

TABLE 203

WET SCRUBBER OPERATING CONDITIONS FOR ROTARY LIME SLUDGE KILN SPECIFICATION

Two sizes of wet scrubbers are to be quoted for each of two efficiency levels. Vendors' quotations should consist of four separate and independent quotations.

	Small	*Large*
Kiln Capacity, ton/day	120	400
Process wt., lb/hr	31,000	103,400
Inlet Gas		
Flow, ACFM	34,000	120,000
Temp., °F	350	350
Flow, SCFM	22,000	77,000
% Moisture (vol)	35	35
Inlet Loading		
Solids Rate, lb/hr	947.4	3,702.6
Solids Loading, gr/DSCF	7.62	8.64
H_2S Rate, lb/hr	9.9	34.5
H_2S Loading, ppm	85	85
Wt.% Soda Fume, as Na_2O	1.8	1.8
Outlet Gas		
Flow, ACFM	28,200	99,600
Temp., °F	168	168
% Moisture (vol)	39	39

Case 1 – Medium Efficiency

Outlet Solids Rate, lb/hr	25.7	40.0
Outlet Solids Loading, gr/DSCF	0.21	0.09
Efficiency, wt.%	97.3	98.9
Scrubber, ΔP	16 in. w.c.	16 in. w.c.

Case 2 – High Efficiency

Outlet Solids Rate, lb/hr	2.42	8.54
Outlet Solids Loading, gr/ACF	0.01	0.01
Outlet Solids Loading, gr/DSCF	<0.015	<0.016
Efficiency, wt.%	99.74	99.77
Wt.% Soda Fume Removal Efficiency	95	95
Scrubber, ΔP	40 in. w.c.	40 in. w.c.

TABLE 204

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR WET SCRUBBERS FOR ROTARY LIME SLUDGE KILNS

	LA Process Wt.		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	34,000	120,000	34,000	120,000
°F	350	350	350	350
SCFM	22,000	77,000	22,000	77,000
Moisture Content, Vol. %	35	35	35	35
Effluent Contaminant Loading				
gr/ACF	3.25	3.60	3.25	3.60
lb/hr	947.4	3,702.6	947.4	3,702.6
Cleaned Gas Flow				
ACFM	28,200	99,600	28,200	99,600
°F	168	168	168	168
SCFM	22,000	77,000	22,000	77,000
Moisture Content, Vol. %	39	39	39	39
Cleaned Gas Contaminant Loading				
gr/ACF	0.11	0.05	0.01	0.01
lb/hr	25.7	40.0	2.42	8.54
Cleaning Efficiency, %	97.3	98.9	99.74	99.77
(1) Gas Cleaning Device Cost	15,627	32,214	18,610	36,997
(2) Auxiliaries Cost				
(a) Fan(s)	12,196	33,623	17,027	45,557
(b) Pump(s)				
(c) Damper(s)				
* (d) Conditioning, Equipment				
* (e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering	45,542	72,037	53,042	83,703
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost	73,365	137,874	88,679	166,257

*Not included.

TABLE 205

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR WET SCRUBBERS FOR ROTARY LIME SLUDGE KILNS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year		8,000	8,000	8,000	8,000
Operating Labor (if any)					
Operator					
Supervisor					
Total Operating Labor					
Maintenance					
Labor	$6/hr				
Materials					
Total Maintenance		2,223	3,312	2,889	4,345
Replacement Parts		1,389	2,404	1,833	9,351
Total Replacement Parts		1,389	2,404	1,833	9,351
Utilities					
Electric Power	$.011/kw-hr	13,386	45,293	22,117	73,705
Fuel					
Water (Process)	$.25/M gal	2,477	9,011	3,435	12,261
Water (Cooling)					
Chemicals, Specify					
Total Utilities		15,863	54,304	25,552	85,966
Total Direct Cost		19,475	60,020	30,274	99,662
Annualized Capital Charges		7,337	13,787	8,868	16,626
Total Annual Cost		26,812	73,807	39,142	116,288

FIGURE 157

CAPITAL COSTS FOR WET SCRUBBERS FOR ROTARY LIME SLUDGE KILNS (MEDIUM EFFICIENCY)

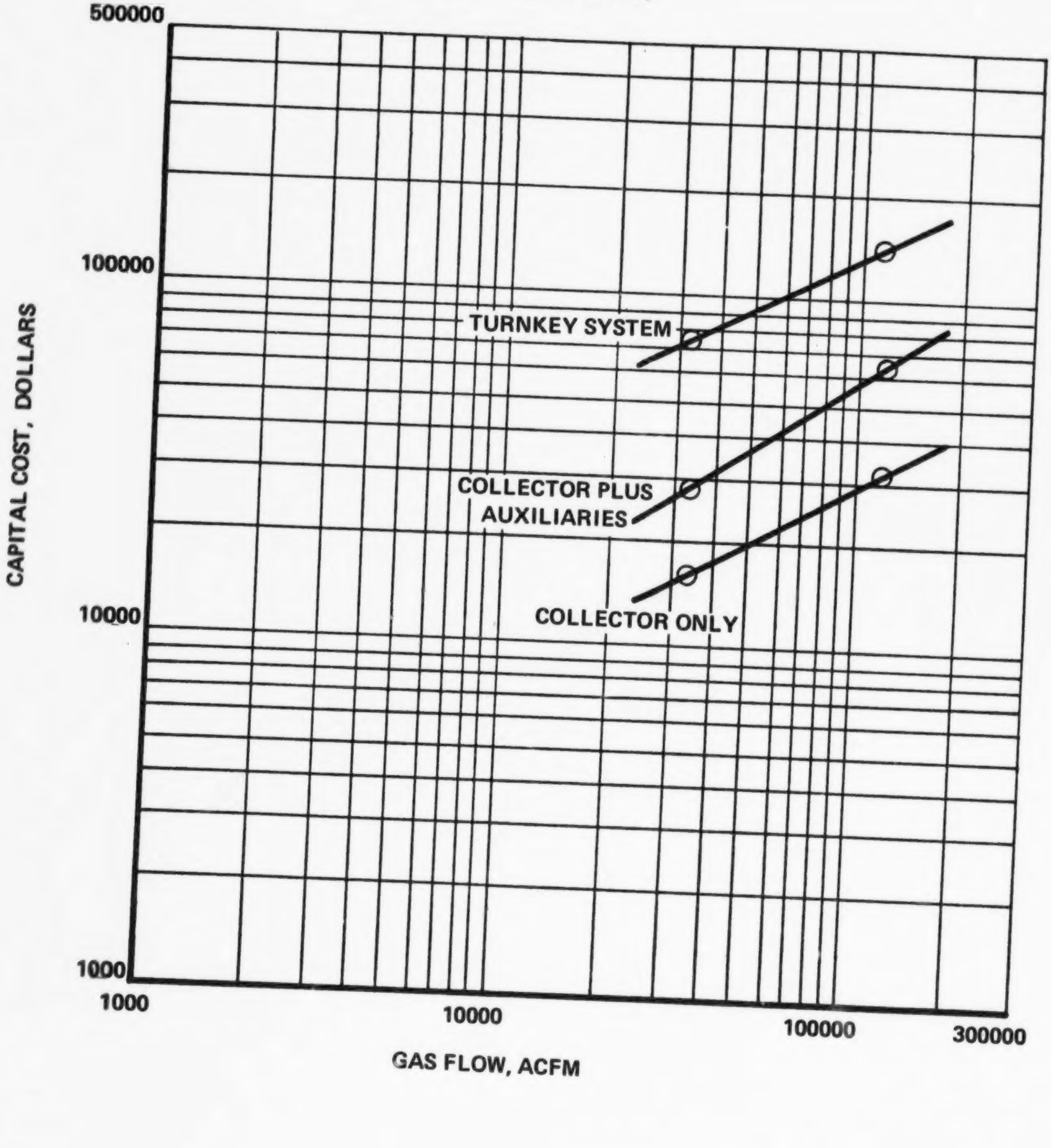


FIGURE 158

ANNUAL COSTS FOR WET SCRUBBERS FOR ROTARY LIME SLUDGE KILNS (MEDIUM EFFICIENCY)

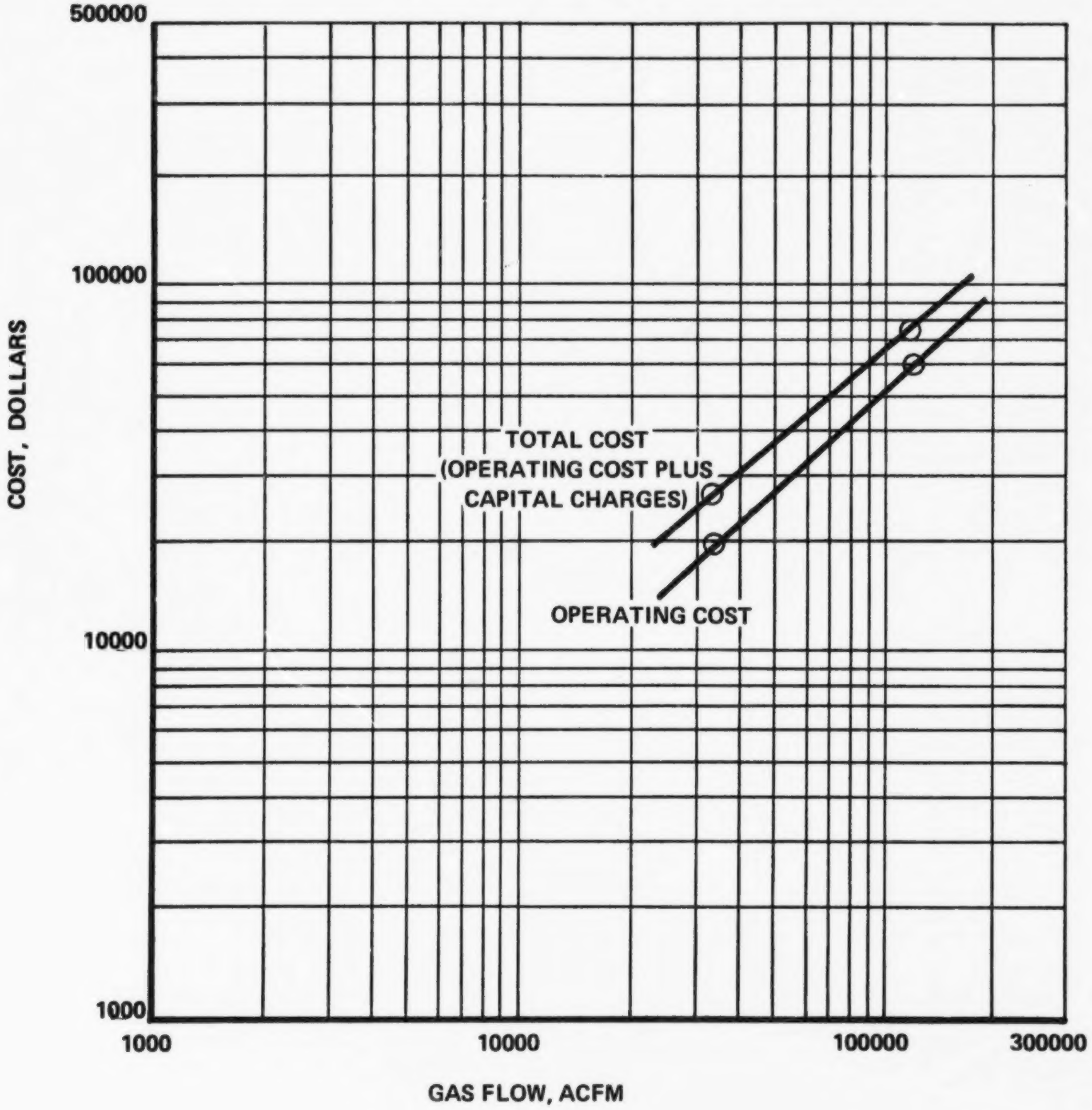


FIGURE 159

CAPITAL COSTS FOR WET SCRUBBERS FOR ROTARY LIME SLUDGE KILNS (HIGH EFFICIENCY)

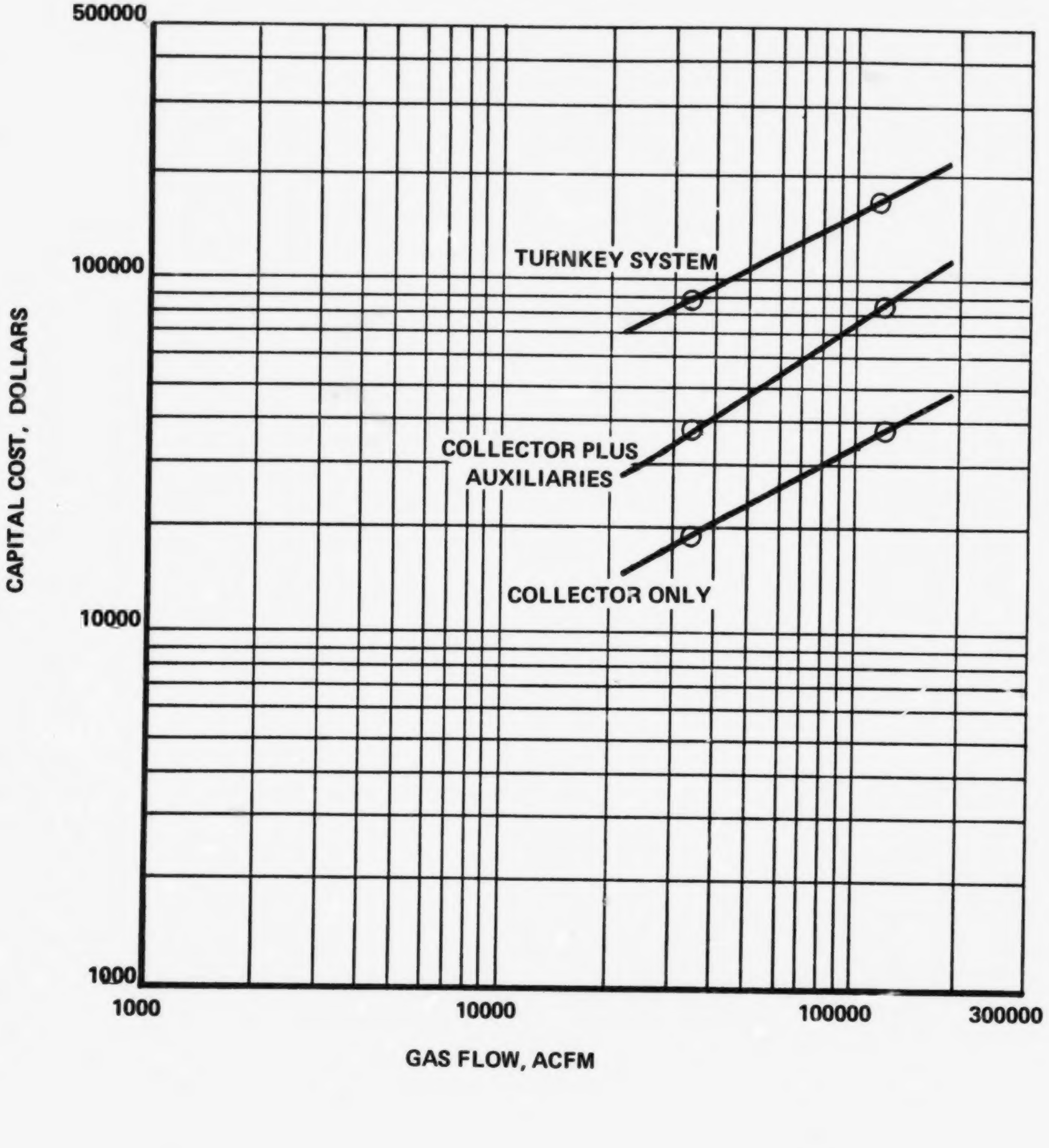


FIGURE 160

ANNUAL COSTS FOR WET SCRUBBERS
FOR ROTARY LIME SLUDGE KILNS
(HIGH EFFICIENCY)

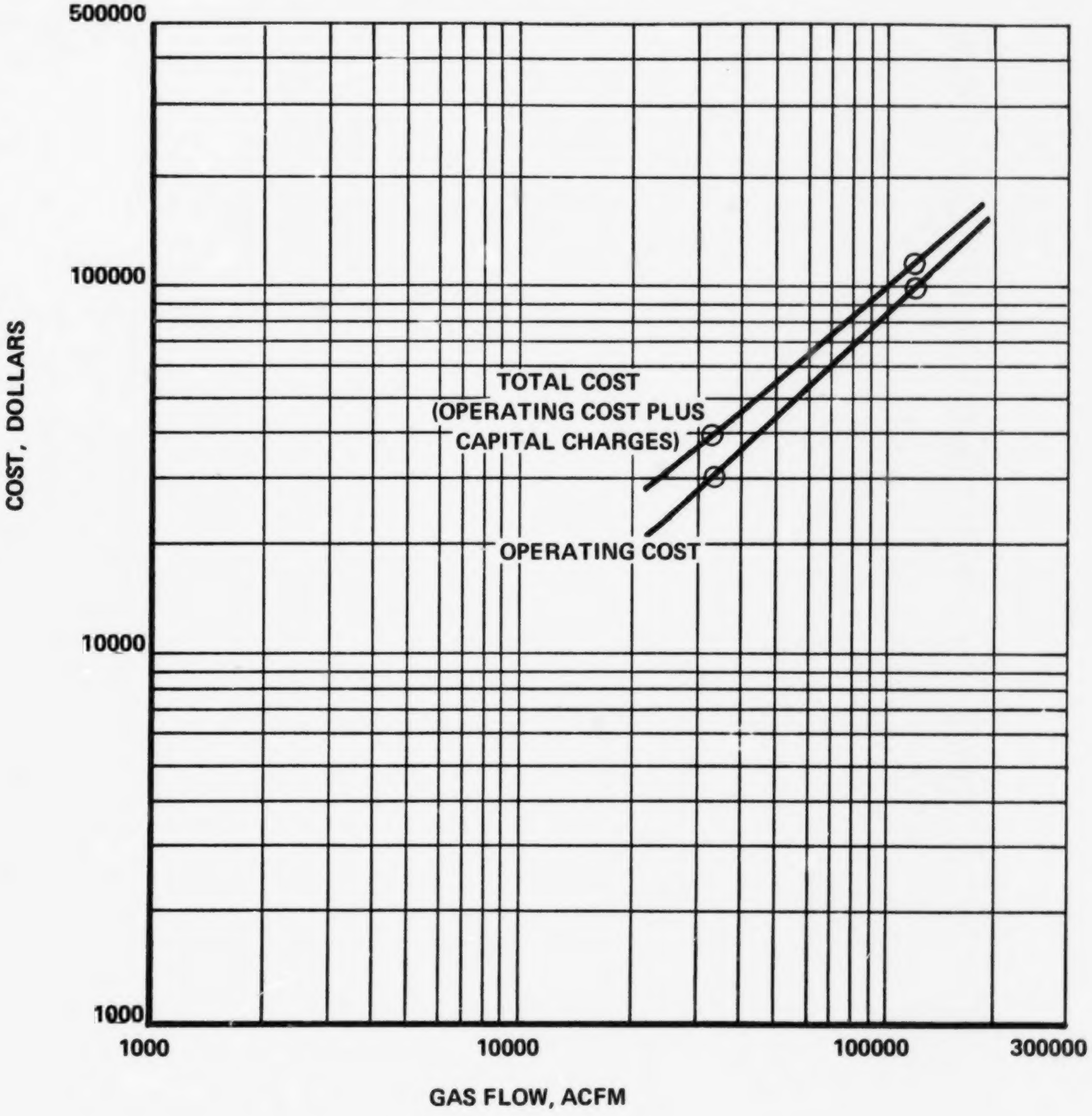


TABLE 206

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR ROTARY LIME SLUDGE KILNS (HIGH EFFICIENCY)

Population Size — 20 Sample Size — 3

Capital Cost = $88,679

Conf. Level, %	Capital Cost, Dollars Lower Limit	 Upper Limit
50	$ 70,015	$107,342
75	52,516	124,841
90	25,590	151,767
95	691	176,666

Capital Cost - $166,257

Conf. Level, %	Capital Cost, Dollars Lower Limit	 Upper Limit
50	$143,333	$189,181
75	121,840	210,674
90	88,768	243,746
95	58,187	274,327

FIGURE 161

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR ROTARY LIME SLUDGE KILNS (HIGH EFFICIENCY)

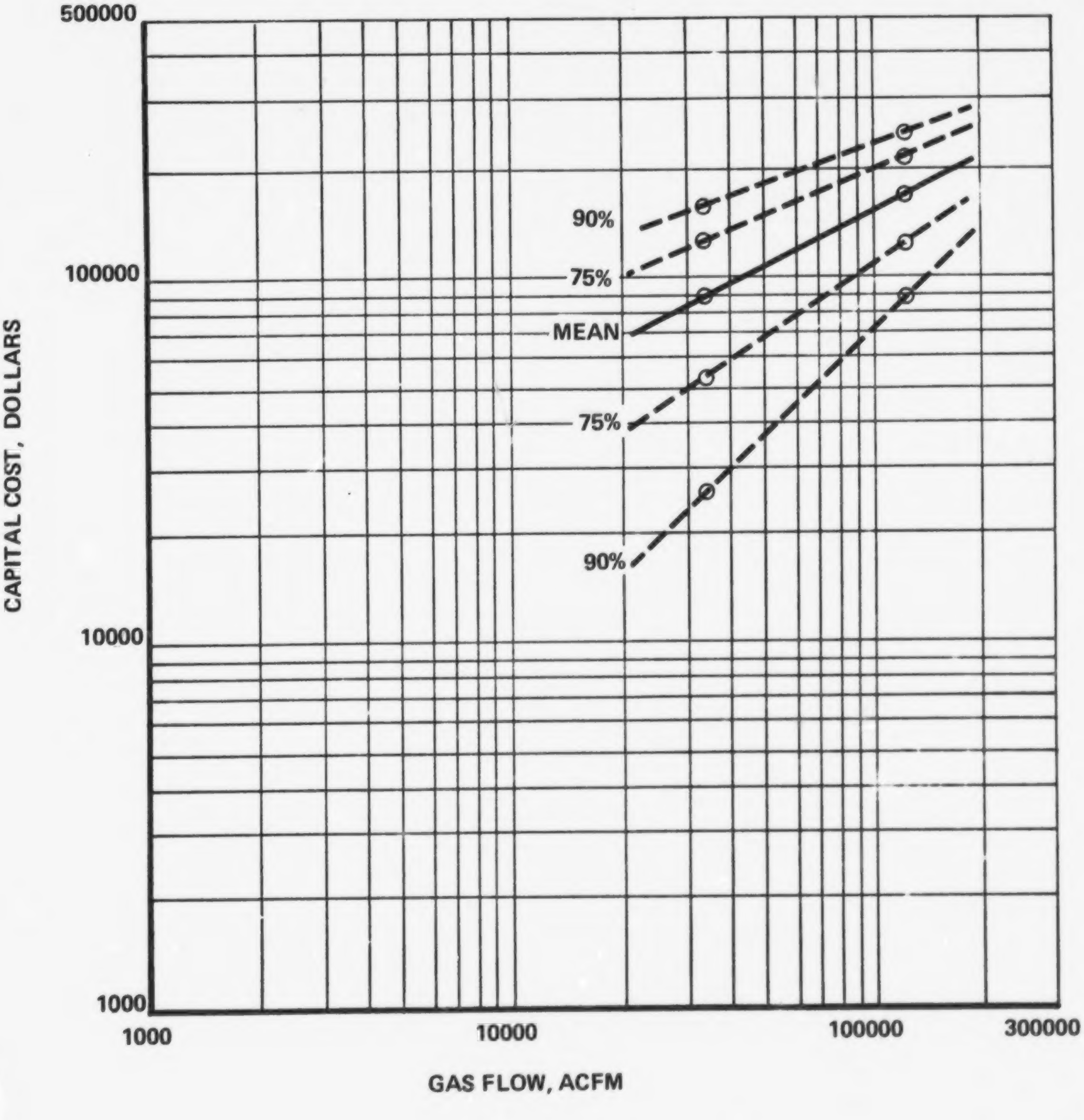


TABLE 207

ELECTROSTATIC PRECIPITATOR PROCESS DESCRIPTION FOR ROTARY LIME SLUDGE KILN SPECIFICATION

A single precipitator is to remove entrained lime dust or soda fume from the exhaust gas of a rotary kiln used to produce reburned lime for the sulfate pulping process. The kiln is fired with natural gas. The kiln is fed with lime sludge containing 55 wt.% solids. If the sludge has been washed properly, it will contain about 0.5 wt.% sodium, expressed as Na_2O *on a dry mud basis. However, this is not always the case, and the sludge to this kiln has a sodium content, expressed as* Na_2O*, of about 2.0 wt.%. The feed end of the kiln is equipped with a dust fallout chamber and heat recuperating chain system.*

The exhaust gas is to be brought from the feed end housing to a location 20 feet to one side of the kiln enclosure and 20 feet above grade. The precipitator will be at ground level in an area beyond the ductwork which is free of space limitations. A fan will exhaust the precipitator outlet gas to a 50 foot high stack. The precipitator is to reduce continuously the kiln outlet loading to the levels specified.

The precipitator system should include the following:

1. *Precipitator provided with a minimum of two independent electrical fields in the direction of gas flow.*
2. *Trough type hoppers equipped with continuous dust removal by screw conveyor to a dust tank. The conveying system must be provided with suitable sealing for negative pressure operation.*
3. *Automatic voltage control.*
4. *Safety interlocked system which prevents access to the interior of the precipitator unless the electrical circuitry is disconnected and grounded.*
5. *Rapping system which is adjustable in terms of both intensity and rapping period.*
6. *Necessary fans and motors.*
7. *Model study for precipitator gas distribution*

TABLE 208

ELECTROSTATIC PRECIPITATOR OPERATING CONDITIONS FOR ROTARY LIME SLUDGE KILN SPECIFICATION

Two sizes of precipitators are to be quoted for each of two efficiency levels. Vendors' quotations should consist of four separate and independent quotations.

	Small	Large
Kiln Capacity, ton/day	120	400
Process wt., lb/hr	31,000	103,400
Inlet Gas		
Flow, ACFM	34,000	120,000
Temp., °F	350	350
Flow, SCFM	22,000	77,000
% Moisture (vol.)	35	35
Dew Point, °F	163	163
Inlet Loading		
Total Solids Rate, lb/hr	947.4	3,702.6
Total Solids Loading, gr/ACF	3.25	3.60
Total Solids Loading, gr/DSCF	7.70	8.55
H_2S, lb/hr	10.0	35.1
H_2S, ppm (vol.)	85	85
Na_2CO_3, lb/hr	14.0	49.5
Na_2CO_3, gr/ACF	0.96	0.96
Na_2SO_4, lb/hr	0.8	2.9
Na_2SO_4, gr/ACF	0.006	0.006
*Case 1 – Medium Efficiency**		
Outlet Solids Rate, lb/hr	25.7	40.0
Outlet Solids Loading, gr/ACF	0.09	0.04
Outlet Solids Loading, gr/DSCF	0.21	0.095
Collection Efficiency, wt.%	97.3	98.9
Drift Velocity, fps	0.25	0.25
*Case 2 – High Efficiency**		
Outlet Solids Rate, lb/hr	2.92	10.28
Outlet Solids Loading, gr/ACF	0.01	0.01
Outlet Solids Loading, gr/DSCF	0.02	0.02
Collection Efficiency, wt.%	99.69	99.72
Wt.% Soda Fume Removal Efficiency	95	95
Drift Velocity, fps	0.25	0.25

*See page 6 for definition of collection efficiency levels.

TABLE 209

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR ELECTROSTATIC PRECIPITATORS FOR ROTARY LIME SLUDGE KILNS

	LA Process Wt.		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	34,000	120,000	34,000	120,000
°F	350	350	350	350
SCFM	22,000	77,000	22,000	77,000
Moisture Content, Vol. %	35	35	35	35
Effluent Contaminant Loading				
gr/ACF	3.25	3.60	3.25	3.60
lb/hr	947.4	3,702.6	947.4	3,702.6
Cleaned Gas Flow				
ACFM	34,000	120,000	34,000	120,000
°F	350	350	350	350
SCFM	22,000	77,000	22,000	77,000
Moisture Content, Vol. %	35	35	35	35
Cleaned Gas Contaminant Loading				
gr/ACF	0.09	0.04	0.01	0.01
lb/hr	25.7	40	2.9	10.2
Cleaning Efficiency, %	97.3	98.9	99.69	99.72
(1) Gas Cleaning Device Cost	67,260	137,405	80,190	158,970
(2) Auxiliaries Cost	25,963	46,153	29,848	48,090
(a) Fan(s)				
(b) Pump(s)				
(c) Damper(s)				
(d) Conditioning, Equipment				
(e) Dust Disposal Equipment				
(3) Installation Cost				
(a) Engineering	60,310	130,228	74,770	142,300
(b) Foundations & Support				
Ductwork				
Stack				
Electrical				
Piping				
Insulation				
Painting				
Supervision				
Startup				
Performance Test				
Other				
(4) Total Cost	153,533	313,786	184,808	349,360

TABLE 210

ANNUAL OPERATING COST DATA (COSTS IN $/YR) FOR ELECTROSTATIC PRECIPITATORS FOR ROTARY LIME SLUDGE KILNS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year					
Operating Labor (if any)	$6/hr				
Operator		2,000	2,000	2,000	2,000
Supervisor					
Total Operating Labor		2,000	2,000	2,000	2,000
Maintenance	$6/hr				
Labor		1,429	5,044	1,429	5,044
Materials					
Total Maintenance		1,429	5,044	1,429	5,044
Replacement Parts		1,000	2,000	1,000	2,000
Total Replacement Parts		1,000	2,000	1,000	2,000
Utilities					
Electric Power	$.011/kw-hr	2,178	8,888	4,796	12,628
Fuel					
Water (Process)					
Water (Cooling)					
Chemicals, Specify					
Total Utilities		2,178	8,888	4,796	12,628
Total Direct Cost		6,607	17,932	9,225	21,672
Annualized Capital Charges		15,353	31,379	18,481	34,936
Total Annual Cost		21,960	49.311	27,706	56,608

FIGURE 162

CAPITAL COSTS FOR ELECTROSTATIC PRECIPITATORS FOR ROTARY LIME SLUDGE KILNS (MEDIUM EFFICIENCY)

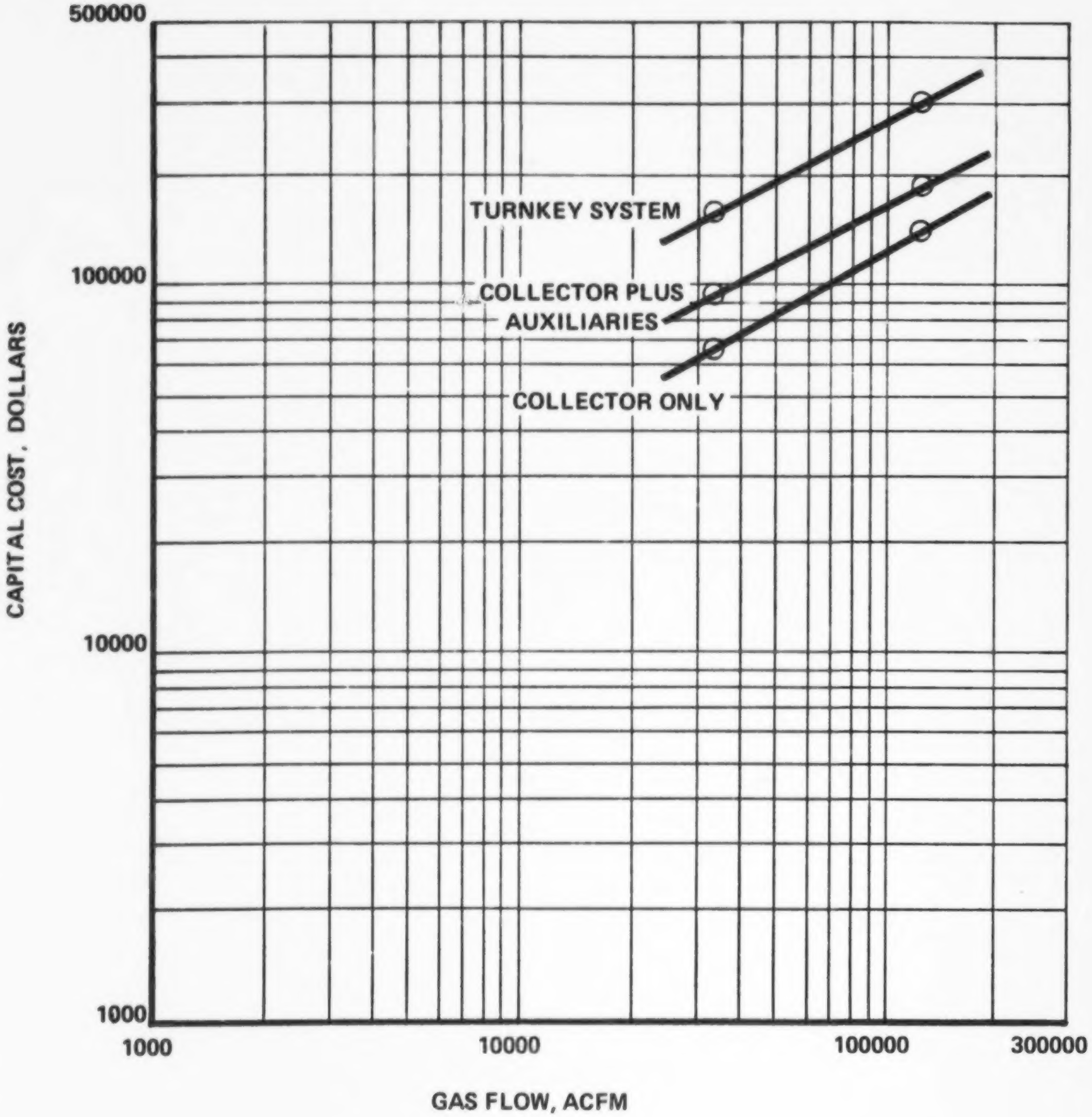


FIGURE 163

ANNUAL COSTS FOR ELECTROSTATIC PRECIPITATORS FOR ROTARY LIME SLUDGE KILNS (MEDIUM EFFICIENCY)

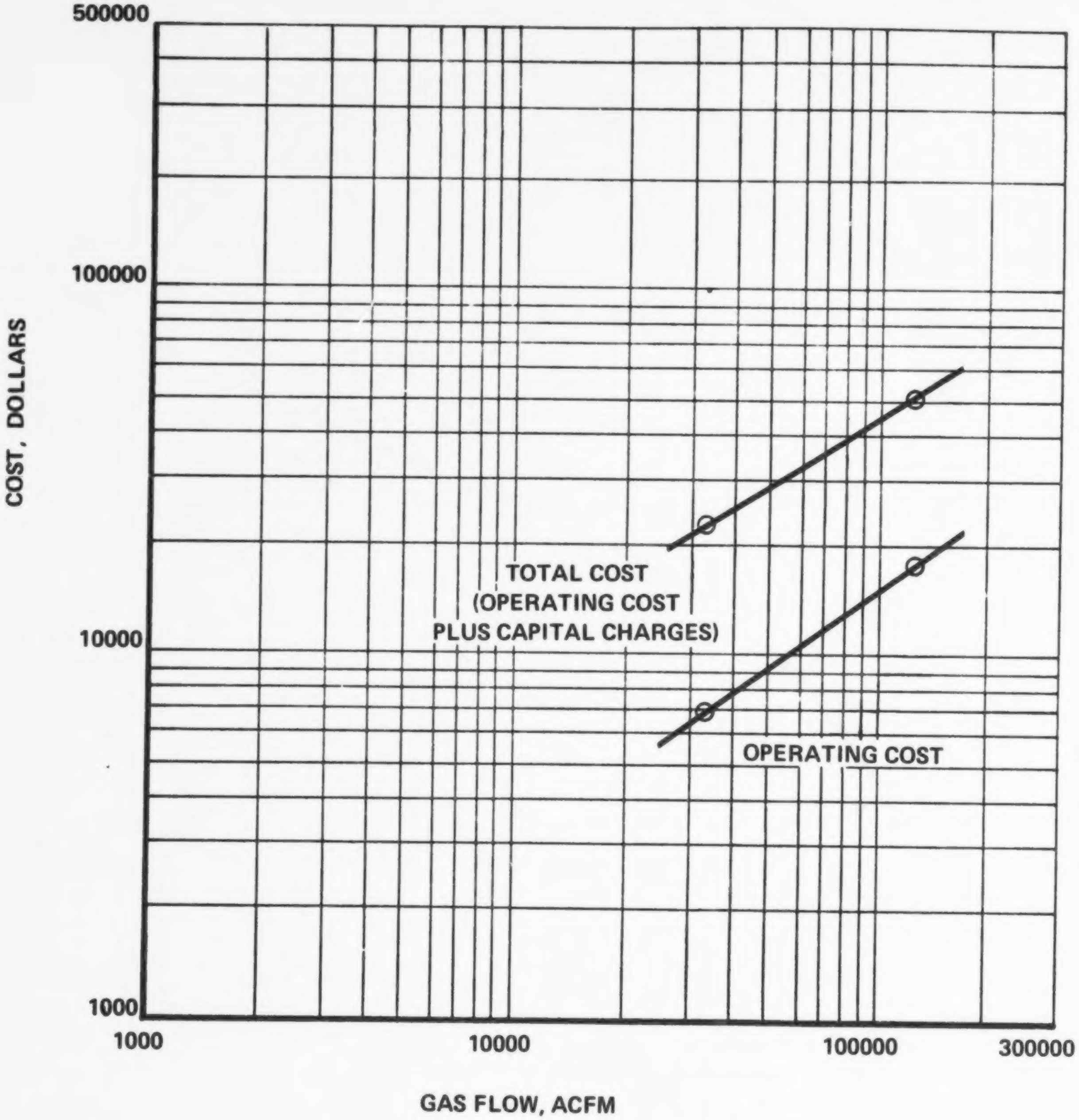


FIGURE 164

CAPITAL COSTS FOR ELECTROSTATIC PRECIPITATORS
FOR ROTARY LIME SLUDGE KILNS
(HIGH EFFICIENCY)

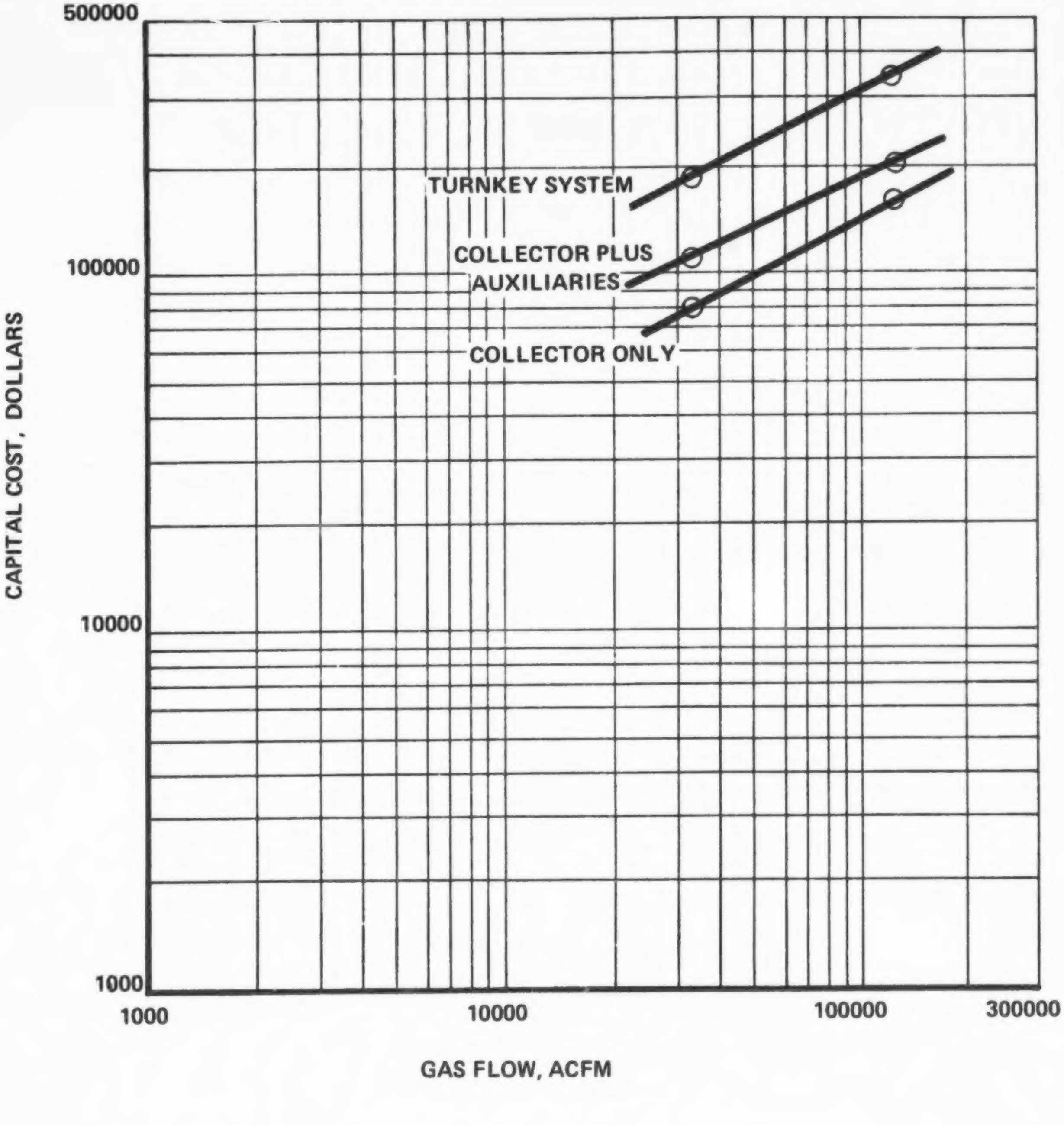


FIGURE 165

ANNUAL COSTS FOR ELECTROSTATIC PRECIPITATORS FOR ROTARY LIME SLUDGE KILNS (HIGH EFFICIENCY)

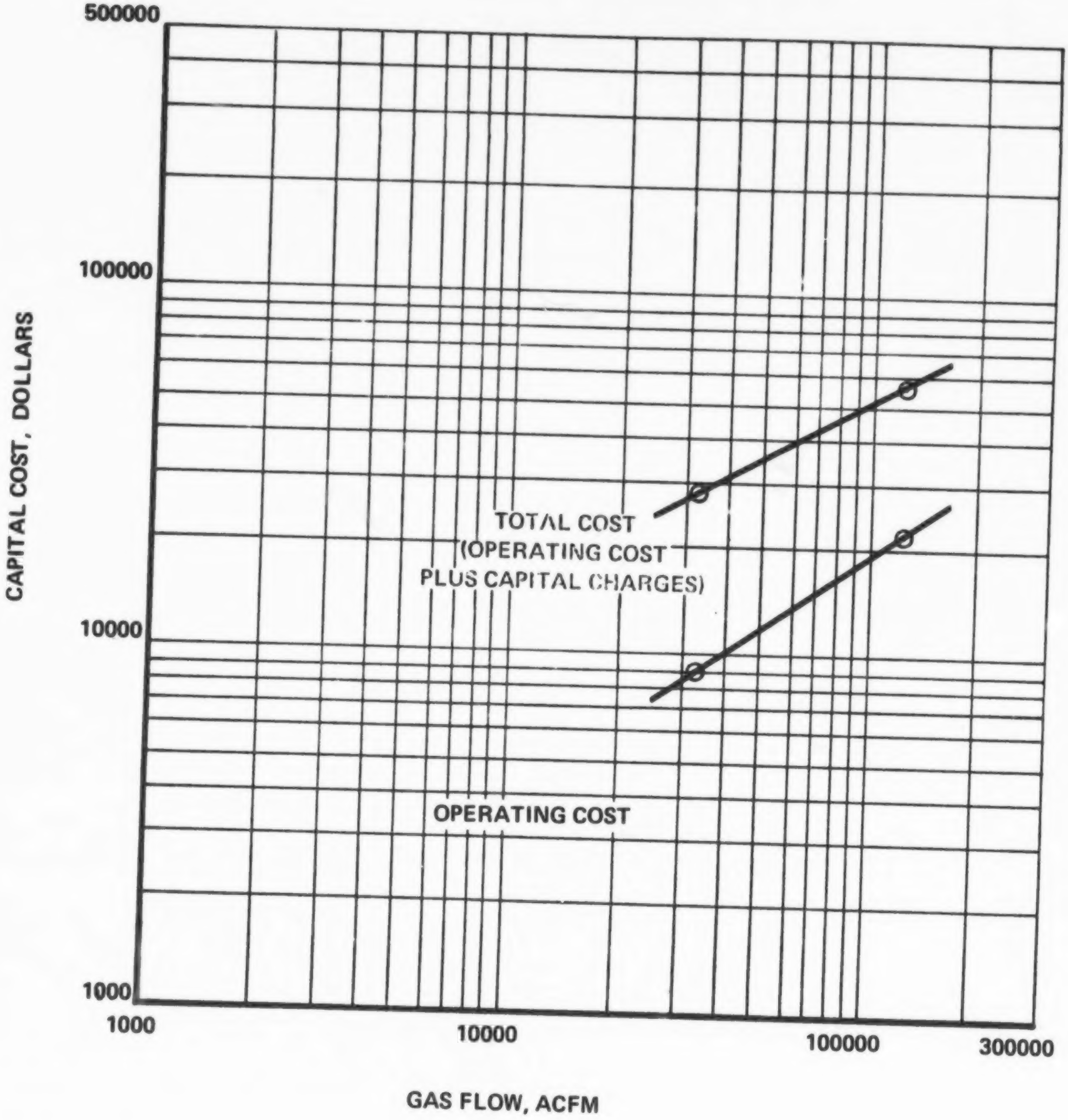


TABLE 211

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF ELECTROSTATIC PRECIPITATORS FOR ROTARY LIME SLUDGE KILNS (HIGH EFFICIENCY)

Population Size – 20 Sample Size – 2

Collector Only Cost = $80,190

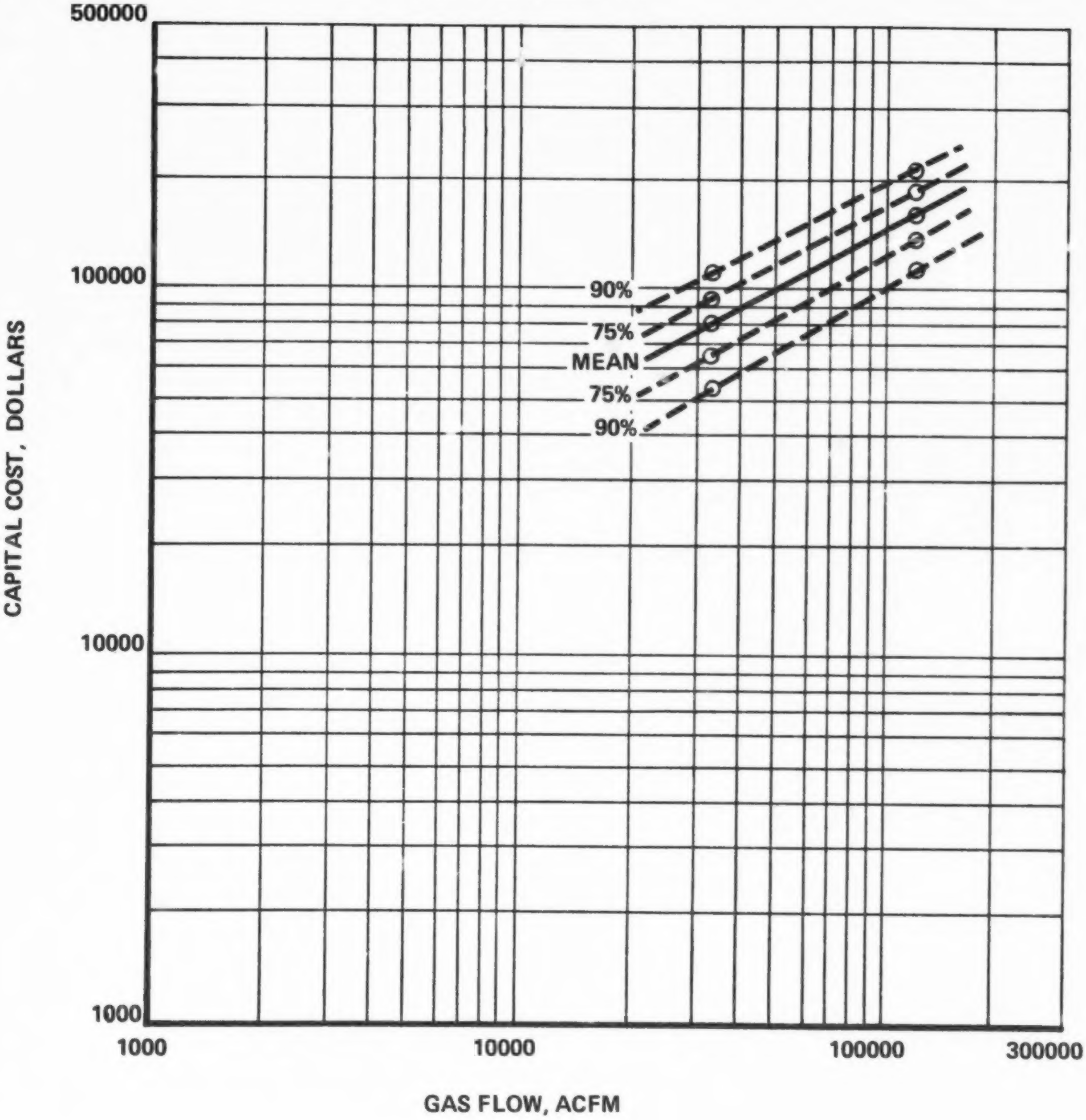
Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$ 74,417	$ 85,964
75	67,404	92,976
90	53,678	106,702
95	38,853	121,527

Collector Only Cost = $158,970

Conf. Level, %	Capital Cost, Dollars Lower Limit	Upper Limit
50	$148,039	$169,901
75	134,763	183,177
90	108,776	209,164
95	80,709	237,231

FIGURE 166

CONFIDENCE LIMITS FOR COLLECTOR ONLY COST OF ELECTROSTATIC PRECIPITATORS FOR ROTARY LIME SLUDGE KILNS (HIGH EFFICIENCY)



REFERENCES

1. Study of Technical & Cost Information for Gas Cleaning Equipment in the Lime & Secondary Non-ferrous Metallurgical Industries, Industrial Gas Cleaning Institute, 1970, pp. 24-61.

2. Lewis, C. J. & Crocker, B. B., "The Lime Industry's Problem of Airborne Dust," J.A.P.C.A., 19(1): 31-39 (Jan. 1969).

3. Boynton, R. S., The Chemistry & Technology of Lime & Limestone, Interscience Publishers, N.Y., 1966.

4. Schwarzkopf, F., "Comparison of Modern Lime Calcining Systems," Rock Products, 73:68-71+ (July, 1971).

5. Stern, A. C., Air Pollution, Vol. III, Academic Press, N.Y., 1968, pp. 243-267.

6. Shreve, R. N., The Chemical Process Industries, McGraw-Hill Book Company, Inc., N.Y., 1956.

7. Casey, J. P., Pulp & Paper, Vol. 1 – Pulp & Bleaching, Interscience Publishers, N.Y., 1960.

8. Britt, K. W., Handbook of Pulp & Paper Technology, Van Nostrand, Reinhold Co., N.Y., 1970.

9. Stephenson, J. N., Pulp & Paper Manufacture, Vol. 1 – Preparation & Treatment of Wood Pulp, McGraw-Hill Book Co., N.Y., 1950.

10. Rydholm, S. A., Pulping Processes, Interscience Publishers, N.Y., 1965.

11. Duprey, R. L., Compilation of Air Pollutant Emission Factors, U. S. DHEW, Public Health Service, Publication No. 999-HP-42, Durham, North Carolina, 1968.

12. Stuart, H. H., & Bailey, R. E., "Performance Study of a Lime Kiln & Scrubber Installation," TAPPI, 48(5):104A-108A (May, 1965).

13. Shah, I. S., "Pulp Plant Pollution Control," Chem. Eng. Progress, 64(9): 66-77 (Sept., 1968).

14. Collins, T. T., Jr., "The Venturi Scrubber on Lime Kiln Stack Gases": TAPPI, 42(1): 9-13 (Jan., 1959).

15. Landry, J. E., and Longwell, D. H., "Advances in Air Pollution Control in the Pulp & Paper Industry," TAPPI, 48(6): 66A-70A (June, 1965).

16. Walker, A. B., & Hall, R. M., "Operating Experience with a Flooded Disc Scrubber . . .," J.A.P.C.A., 18(5): 319-23 (May, 1968).

17. Walther, J. E., and Amberg, H. R., "A Positive Air Quality Control Program at a New Kraft Mill," J.A.P.C.A.; 20(1): 9-18 (Jan., 1970).

7. GRAY IRON FOUNDRIES

Iron foundries are shops where iron and steel are melted and cast. These shops exist over a wide range of sizes from small family businesses to large shops owned by giant corporations. The total foundry population has shown a steady decline since the end of World War II from 3,200 in 1947 to 1,600 in 1971 mostly due to closing small shops. The size of the average foundry has risen from an annual production of 3,800 to 8,700 tons(1) during the same period. These figures show that while small foundries are closing down, their production is being more than offset by new large foundries, resulting in a slow but positive growth in total foundry capacity. The distribution of foundry sizes in 1969 is shown in Figure 167.(2)

Small foundries are generally defined as those which employ less than 50 people. They are most often "jobbing" foundries which make small quantities or job lots of a variety of cast products for independent customers. Large foundries are often described as *captive* in that they are owned by their principal customer. They tend to produce large quantities of a few specialized castings. Examples of industries which rely mainly on captive foundries are automobile manufacturers and cast iron pipe producers.

Iron foundries produce three forms of cast iron: gray iron, ductile iron, and malleable iron. Gray iron is produced in the greatest amount as indicated on Table 212.(1) All three forms of cast iron are alloys composed principally of iron and carbon.

All three contain carbon in the graphitic form. They differ from one another by the size and shape of the graphite particles dispersed throughout the alloy. The graphite in gray iron appears in large crinkled flakes. The flakes have no strength and cause the casting to be brittle. When gray iron castings are broken, the fracture goes through the graphite flakes, producing a gray color in the fracture. It is from this property that gray iron gets its name. Gray iron cast in such a way that all of the carbon occurs as the iron-carbon compound cementite (Fe_3C) is called white iron. Fractures in white iron also occur along the carbon containing phase. Cementite, however, is steely white in appearance and the fractures, therefore, are light colored. Iron cast with high carbon and silicon contents will form gray iron. Lower carbon and silicon contents will form white iron.(4)

The cementite phase of white iron can be converted into graphite by heating in the presence of sufficient silicon after the casting has solidified. The graphite formed by this process is different from the flakes which are present in

Preceding page blank

FIGURE 167

DISTRIBUTION OF IRON FOUNDRY SIZES (1969)

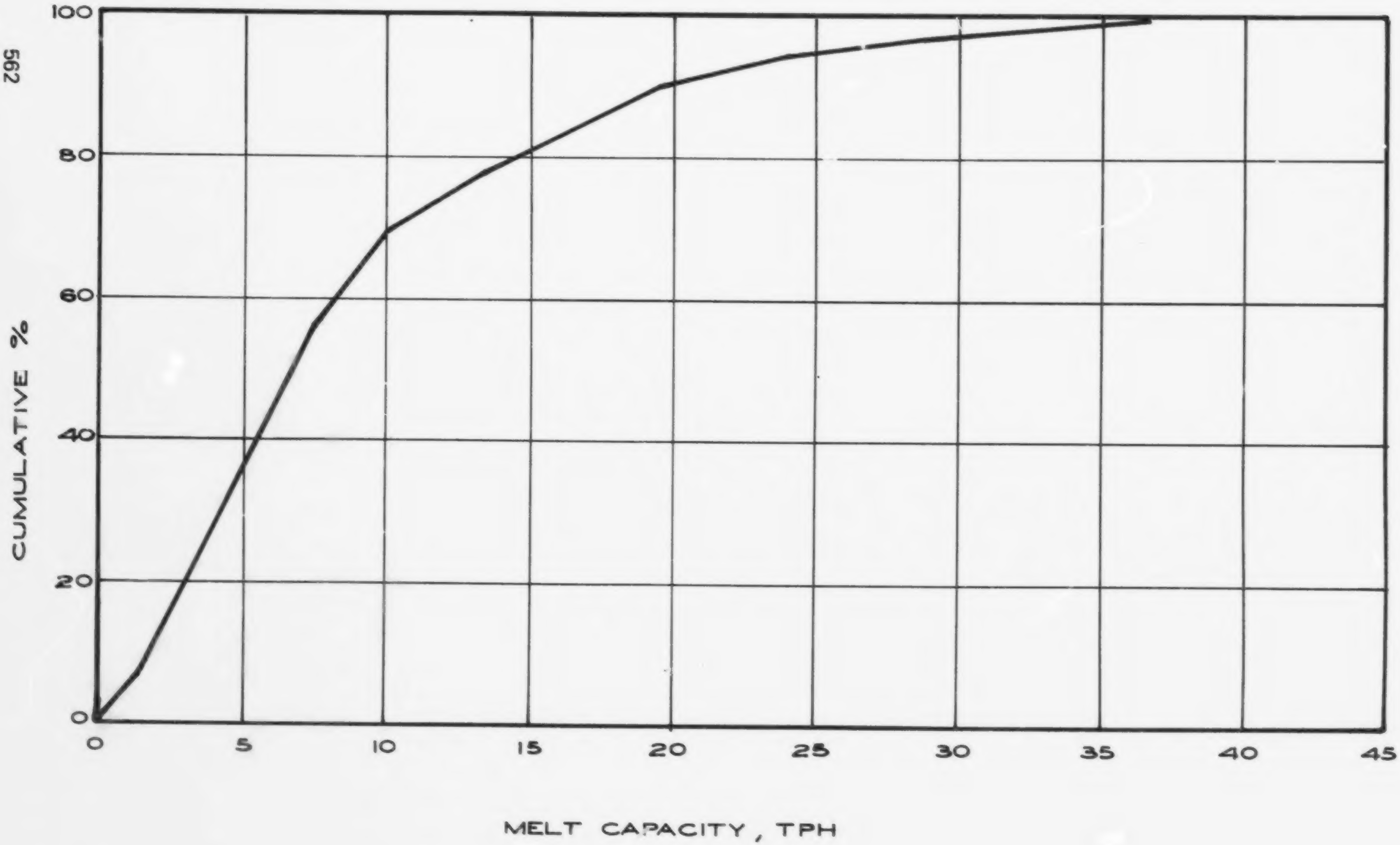


TABLE 212

AVERAGE ANNUAL FOUNDRY PRODUCTION IN THE UNITED STATES[1]

Type of Iron	Average Production ton/yr*	Percent of Total Production
Gray Iron	11,650,000	84.3
Malleable Iron	1,075,000	7.8
Ductile Iron	1,092,000	7.9
Total	13,817,000	100.0

*For the five year period 1965 through 1969.

gray iron. The graphite is formed in spheroids which do not embrittle the cast metal as much as the graphite flakes. As a result some deformation of the casting is possible without fracture. Metal produced in this way is called malleable iron. A comparison of the physical properties of gray iron and malleable iron is presented in Table 213.

Further alteration in the form of the graphite can be made by innoculating the melt with a small amount of magnesium just before casting. The result is a cast iron containing very small spherical nodules of graphite. It is called ductile or nodular iron. It has ductility equivalent to malleable iron and has greater tensile strength. Its properties are also shown in Table 213.

Differences in the production of these three forms of cast iron are illustrated in Figure 168.(1) All of the operations shown apply to all three kinds of cast iron unless noted otherwise on the drawing.

FOUNDRY OPERATIONS

Foundries engage in a great many separate operations in order to produce iron castings. The major operations include:

1. Raw material storage
2. Iron melting
3. Pouring
4. Mold making
5. Core making
6. Shakeout
7. Finished product handling
8. Sand reclamation

Figure 169 shows a diagram of the flow through a foundry and relationships among these eight operations. Most foundries carry out all eight operations, and all have at least the first three in the list. Most of the eight operations involve emission sources and these, too, are indicated on the flow diagram.

TABLE 213

PROPERTIES OF CAST IRONS

	Gray Iron	Malleable Iron	Ductile Iron
Chemical Composition[1]			
Carbon, wt.%	3.00 to 3.75	2.00 to 2.65	3.00 to 3.55
Silicon, wt.%	1.10 to 2.80	0.90 to 1.65	2.25 to 2.55
Manganese, wt.%	0.5	0.25 to 1.25	0.50
Magnesium, wt.%	–	–	0.04 to 0.1
Physical Properties[1],[4]			
Tensile Strength, psi	20,000 to 51,000	50,000 to 77,000	75,000 to 120,000
Yield Strength, psi	NA	32,000 to 70,000	28,000 to 45,000
Brinell Hardness	160 to 270	110 to 245	120 to 270
Elongation, %	NA	3.5 to 22	1 to 15
Impact Strength, ft-lb	1 to 3	4 to 16	NA

FIGURE 168

SCHEMATIC FLOW DIAGRAM
OF CAST IRON PRODUCTION

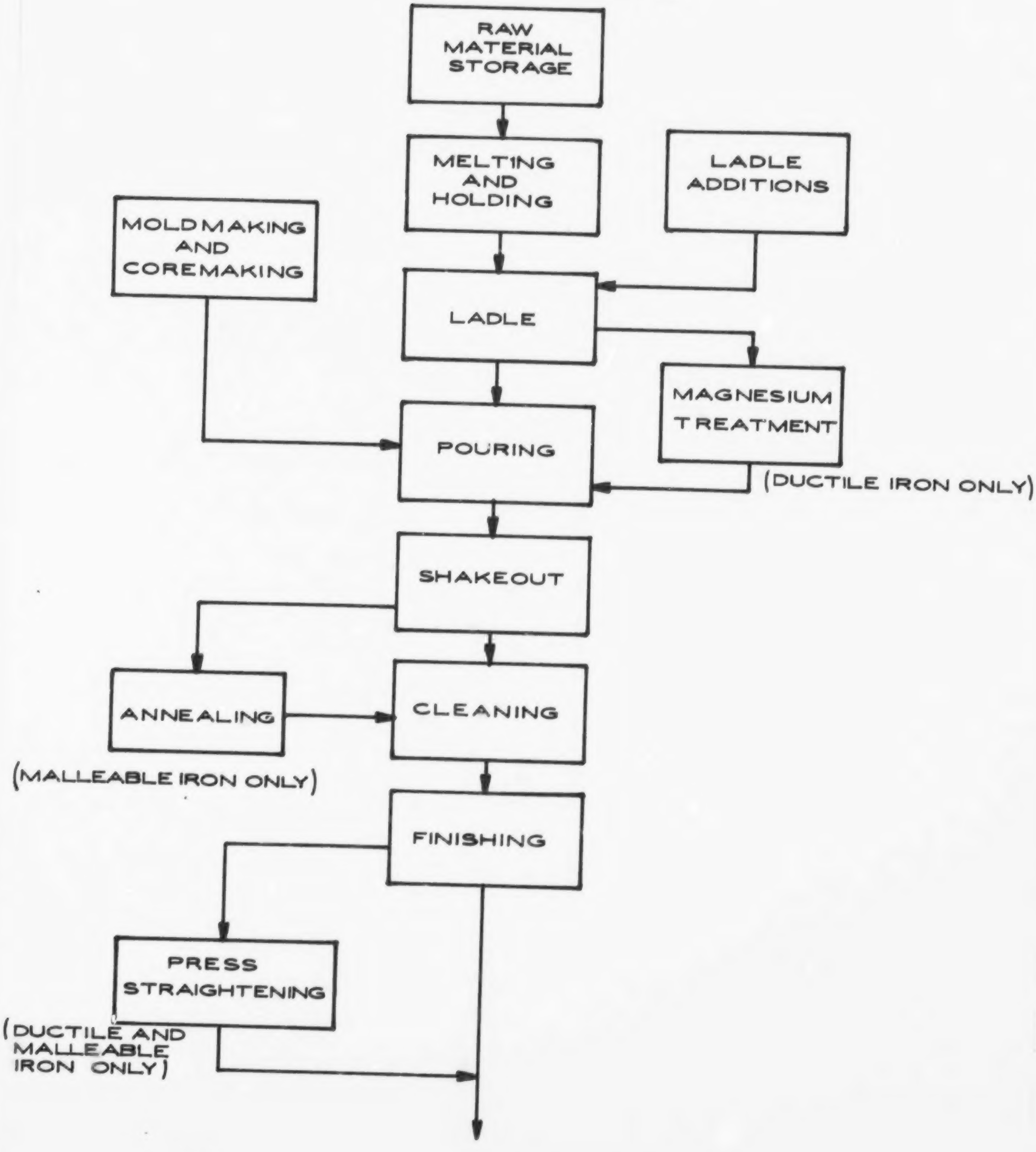
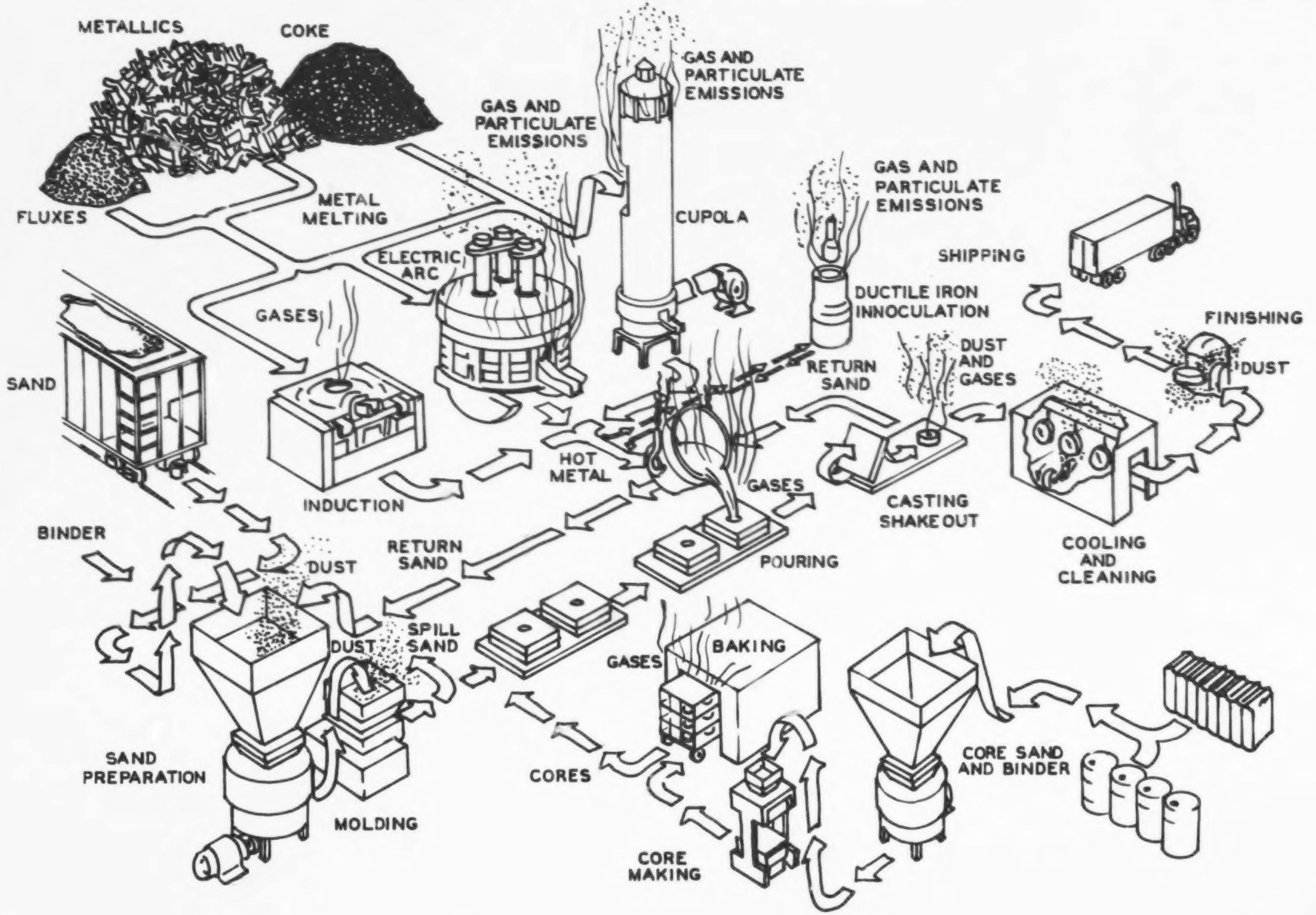

RAW MATERIAL STORAGE
MELTING AND HOLDING
LADLE ADDITIONS
MOLDMAKING AND COREMAKING
LADLE
MAGNESIUM TREATMENT
(DUCTILE IRON ONLY)
POURING
SHAKEOUT
ANNEALING
(MALLEABLE IRON ONLY)
CLEANING
FINISHING
PRESS STRAIGHTENING
(DUCTILE AND MALLEABLE IRON ONLY)

FIGURE 169

FLOW DIAGRAM OF MAJOR IRON FOUNDRY OPERATIONS


METALLICS
COKE
GAS AND PARTICULATE EMISSIONS
GAS AND PARTICULATE EMISSIONS
FLUXES
METAL MELTING
CUPOLA
GAS AND PARTICULATE EMISSIONS
ELECTRIC ARC
SHIPPING
GASES
DUCTILE IRON INNOCULATION
FINISHING
SAND
RETURN SAND
DUST AND GASES
DUST
HOT METAL
GASES
INDUCTION
CASTING SHAKE OUT
BINDER
DUST
RETURN SAND
POURING
COOLING AND CLEANING
SPILL SAND
DUST
GASES
BAKING
SAND PREPARATION
CORES
CORE SAND AND BINDER
MOLDING
CORE MAKING

Raw materials used in the production of iron castings are iron from various sources, alloying agents, fluxes, sand, and fuel. The iron is melted in a furnace and brought to the proper composition with appropriate alloying agents. The molten metal is then tapped into a ladle and poured into a sand mold. The mold is cooled, solidifying the metal. The mold is removed from the casting in the shakeout area and the casting is sent to cleaning and finishing.

Molds and cores are prepared in many ways, depending upon the nature of the product. Each mold making and core making process has its own recipe and formulation. In general, sand is mixed with water and a binding agent, and then formed with a pattern into the desired shape. The mold is dried or heated to fix the shape prior to pouring. After pouring and cooling, the mold is broken and the casting removed. This may be done manually or mechanically on a large vibrating screen. Where possible, the sand from shakeout is recovered and reclaimed by crushing followed by abrasive or thermal removal of spent binding agent. The reclaimed sand is sent back to mold making for use in new molds.

Four types of furnaces are used for iron melting in foundries: the cupola, electric arc, induction, and reverberatory furnaces. Cupolas are the most numerous but their position is declining relative to electric arc and induction furnaces. This decline is occurring because cupolas have much higher emission rates than do the other types. Figure 170 shows the change in the numerical distribution of furnace types since 1959.(1)

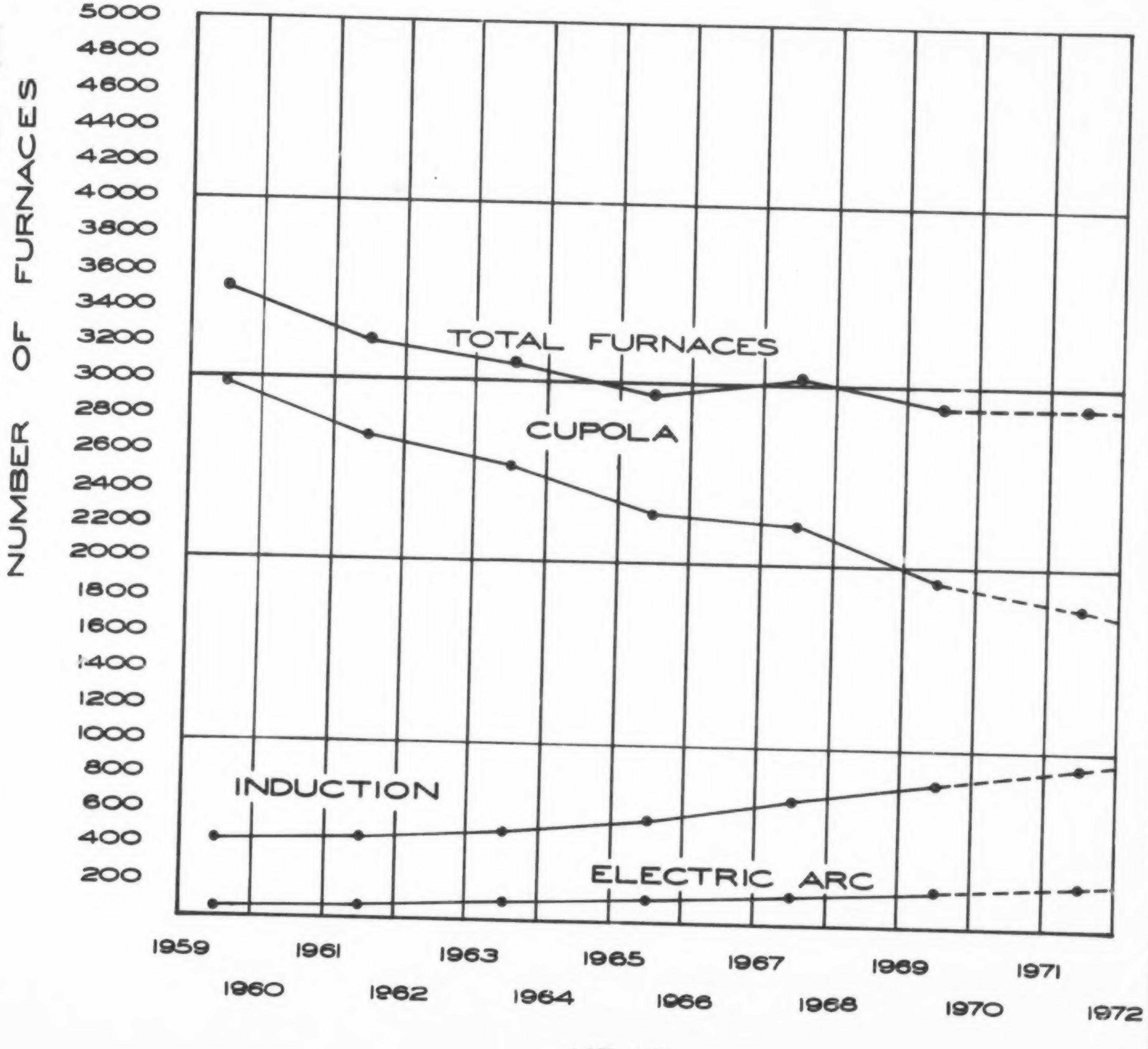
Cupolas are vertical cylindrical furnaces which are charged with alternate layers of iron, fuel (usually coke), and flux. The energy for melting the iron comes from the combustion of the coke. Combustion air is forced through the bed from the bottom with a blower. In some cupolas natural gas is added with the combustion air as a supplementary fuel, and it is also fairly common practice to enrich the gas stream with oxygen. Preheating of the combustion air by direct-fired gas burners may also be used. Slag and molten iron are withdrawn through separate tapholes at the bottom or in many cases the cupola has only one taphole for withdrawal of, first the iron and then the slag.

Electric arc furnaces are cup-shaped vessels with roofs. Three carbon electrodes protrude through the roof and extend into the metal charge held by the furnace. The energy required for melting is obtained from the arc.

Induction furnaces are also cup- or drum-shaped vessels that use electrical

FIGURE 170

TRENDS IN TYPES OF IRON FOUNDRY FURNACES(1)



energy to melt metal. Induction furnaces operate on the same principle as transformers. A primary coil is energized with alternating current, which sets up a magnetic field around it. The magnetic field induces eddy currents in the charge which are converted into heat by the electrical resistance of the charge.

The use of reverberatory air furnaces is expected to decline as they are replaced by electric furnaces.

The melting operation causes the majority of the emissions from a foundry. Although all types of melting furnaces have significant emissions, the cupola is by far the most difficult to properly control. The remainder of this section will be limited to a discussion of control of emissions from cupolas.

THE GRAY IRON CUPOLA

Although its use is steadily declining, the cupola is the work-horse of the foundry industry. Figure 5 shows a diagram of a typical cupola. The furnace is a vertical cylinder charged with alternate layers of iron, coke, and flux.

A layer of sand is generally packed into the bottom of the cupola prior to start up and stays under the iron, coke and flux bed during operation. Air ports, or tuyeres, are located at the bottom. Blast air is forced into the furnace through these tuyeres using a forced draft blower.

Air is required for two purposes. The greater portion is used as combustion air for the coke. This combustion generates the energy for melting the iron. Additional air is required to combust the carbon, silicon, and manganese which may be present in the metal to be melted.(3) The amount of these materials left unoxidized in the molten iron plays a large role in determining the properties of the cast iron.

The air blown into a cupola may be preheated to increase the thermal efficiency of the system. Two methods of preheating are available:

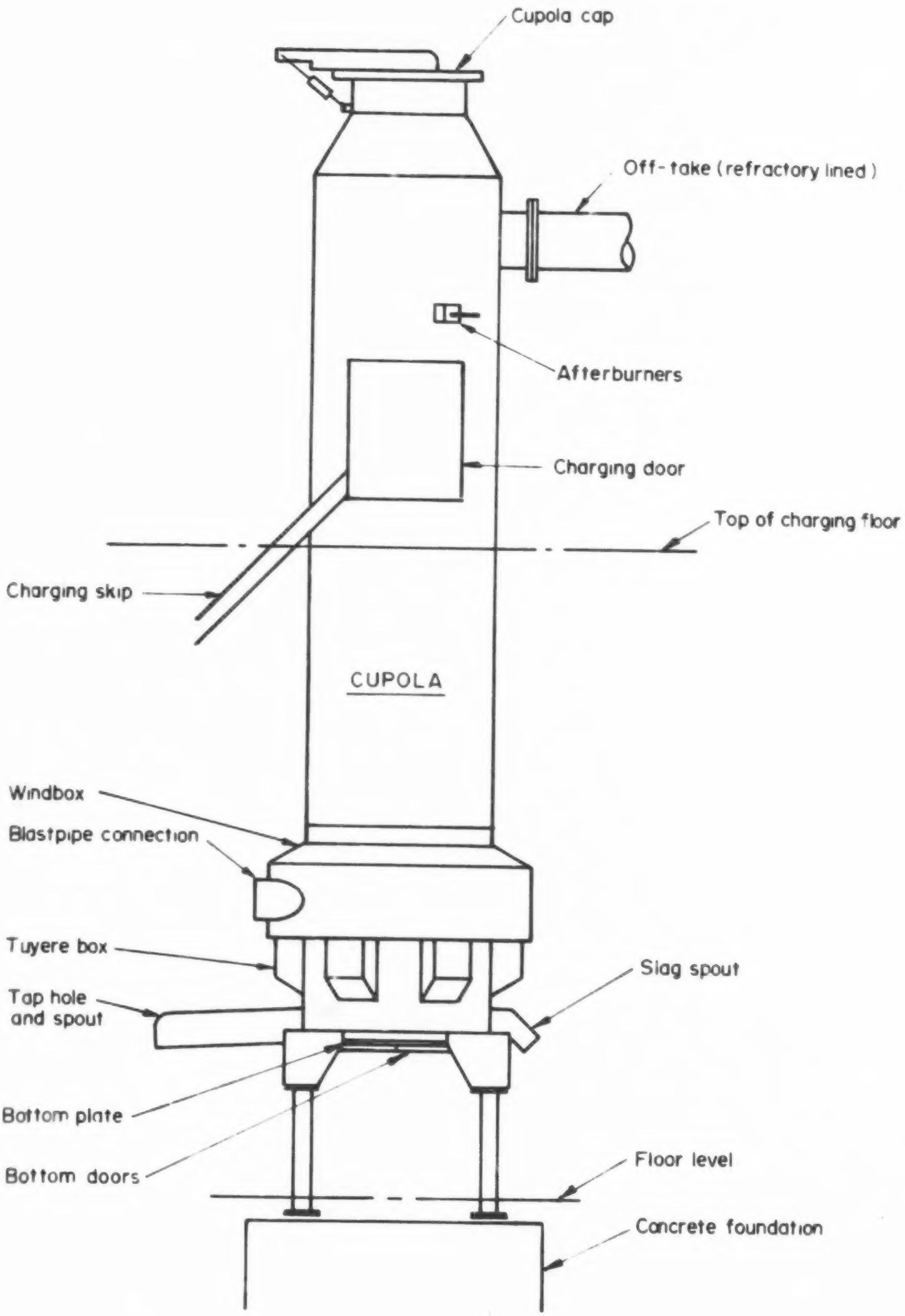
1. A separate external preheater for the combustion air.

2. Combustion of the CO in the exhaust gases to CO_2 followed by heat exchange with the combustion air.

There is a current EPA demonstration project on recuperative heat exchange for an iron foundry cupola,(5) however, the method is not used conventionally.

FIGURE 171

TYPICAL GRAY IRON CUPOLA


Cupola cap
Off-take (refractory lined)
Afterburners
Charging door
Top of charging floor
Charging skip
CUPOLA
Windbox
Blastpipe connection
Tuyere box
Slag spout
Tap hole and spout
Bottom plate
Bottom doors
Floor level
Concrete foundation

The methods involving heat recovery are not in common use at the present time due to high capital cost, high maintenance cost, and low operating rate of many foundries.

Raw materials are charged to the furnace through a door located in the side of the cupola above the charge bed. Small cupolas are charged by hand. Larger units use mechanical charging equipment such as skip hoists.

Exhaust gases are removed from the cupola either from the top, as indicated on Figure 171, or through an opening in the side, immediately below the charging door.

Operation of the cupola is begun by putting coke into the bottom and igniting it. Additional charge materials are mixed into a charge bin and conveyed into the cupola. As the metal melts and the coke is consumed, the entire contents of the furnace shift downward and are replaced by new charge material. The melting is generally considered to occur in three zones. As the new charge moves downward in the cupola, it is preheated by the rising hot combustion gases. These gases include carbon dioxide, carbon monoxide, nitrogen, hydrogen, and sulfur dioxide. There is no free oxygen. As the hot metal enters the second zone, called the melting zone, the atmosphere is highly reducing. As the molten metal continues downward, it enters the third or combustion zone where it is superheated to the desired tapping temperature. The atmosphere in the combustion zone is highly oxidizing.

The gas flow through the cupola is countercurrent to the metal flow. Blast air enters through the tuyeres into the combustion zone. In this zone it combines with the coke as follows:

$$O_2 + C \rightleftharpoons CO_2 + \text{heat}$$

The oxidation of impurities in the metal charge also occurs in this zone. For example:

$$Si + O_2 \rightarrow SiO_2$$

Some oxidation of iron also occurs.

As the gas moves upward, the oxygen content is depleted and the following reaction begins to occur to a greater extent:

$$CO_2 + C \rightleftharpoons 2CO \; 1 \; \text{heat}$$

Water vapor is also reduced in this zone producing carbon monoxide and hydrogen as follows:

$$C + H_2O \rightarrow CO + H_2 - \text{heat}$$

The maximum concentration of CO_2 (14 to 18%) and maximum temperature (2,800 to 3,400° F) occur at the boundary of the combustion and melting zones.(1) As the gases enter the preheating zone, they give up heat to the new charge and their temperature is lowered sufficiently to prevent further reactions from occuring between carbon, oxygen, and oxides of carbon.

Nature of the Gaseous Discharge

The gases which leave a cupola consist of the combustion products from the cupola reactions. The principal components are: nitrogen, carbon dioxide, oxygen, carbon monoxide, hydrogen, and sulfur dioxide.

The oxides of carbon result from the coke as described in the previous section. Sulfur dioxide results from the combustion of the sulfur contained in the coke. The nitrogen is brought in with the combustion air. Oxygen appears primarily as a result of air coming in through the charging door. The relative proportion of these components is highly variable. It depends primarily upon the rate of flow and temperature of blast air, the amount of air coming through the charge door, the cupola operating temperature, the ratio of coke to iron in the cupola, and the composition of the coke. Table 214 lists some examples which demonstrate the variability.(1)

The rate of flow of exhaust gas is also variable and depends on many of the same variables which affect the composition of the gas. It is most strongly influenced by the quantity of blast air and the quantity of air which enters through the charge door. A good average value is 1,000 SCFM per ton/hr of melt capacity. The off-gas temperature can vary from 500 to 2,200° F depending upon: whether a CO afterburner is used, blast air rate, preheat temperature, and the rate of air entering through the charge door.

The gas discharge temperature is cyclic, and decreases as cold material is charged into the cupola. The temperature rises gradually as the cold material warms up. This short cycle, measured in minutes, is superimposed on a longer cycle of light-off, melting-pouring, and burndown. After a period of a few hours to a few days of melting, it is necessary to shut the cupola down for

TABLE 214

EXHAUST GASES COMPOSITIONS FROM FOUR DIFFERENT CUPOLAS(1)

	Mole Percent in Sample			
Component	A	B	C	D
CO	1	3	2	0
CO_2	3	16	4	13
O_2	17	1	7	7
N_2	75	80	86	80
H_2	3	–	–	0
	99	100	99	100

relining. The resulting "burndown" produces the highest emission temperatures during the operating cycle, because no cold charge is being added during burndown. Frequently, two cupolas are operated in tandem, so that one can melt while another is relined. A single gas cleaning system is used for the pair.

Table 215 shows a weight balance for a theoretical cupola melting one ton/hr of iron. The theoretical cupola has the following characteristics:

1. Lined construction
2. Gas takeoff above the charge door
3. Charge door open during operation
4. Preheated blast air, unenriched

Pollution Control Considerations

The control system which serves a cupola must deal with at least two distinct types of emissions: carbon monoxide and particulate matter. Emission control systems which deal with these two problems consist of three parts:

1. An afterburner (combustion chamber or upper cupola stack) to convert CO to CO_2 and to burn organic matter
2. A cooling device to lower the afterburner exit temperature to a level suitable for conventional abatement equipment
3. A collector to remove particulate matter

The afterburner may consist of a stack above the charge door containing gas burners or torches. These should be designed to raise the exit gas temperature to at least 1,200°F.(3) The geometry of the installation should provide a minimum of 1/4 sec. of residence time. In large cupolas it is necessary to create sufficient turbulence to insure that the combustibles and air become properly mixed. This may alter the residence time required. Stratification of the gas stream in large diameter cupolas can make mixing a problem.

TABLE 215

MATERIAL BALANCE FOR THEORETICAL CUPOLA(7)

Input Material	lb/hr	wt.%
Pig Iron	97	2.17
Silvery Piglets	66	1.48
Purchased Scrap	873	19.51
Steel Scrap	388	8.67
Returns	582	13.01
Total Metal Charge	2006	44.84
Coke	268	5.99
Natural Gas	0	–
Fuel Oil	0	–
Flux and Additives	40	0.89
Air	2139	47.81
Oxygen	0	–
Cupola Lining	21	0.47
Total	4474	100
Output Material		
Molten Iron	2000	44.95
Slag	53	1.19
Emission Dust	20	.45
Nitrogen	1636	36.77
Carbon Dioxide	489	10.99
Carbon Monoxide	248	5.57
Hydrogen	2	.04
Sulfur Dioxide	1	.02
Total Off Gases	2376	
Total	4449	100

Top Gas Molecular Weight – 31.24
Top Gas Rate, SCFM – 461
Dust Concentration, gr/SCF – 5.07
Air in Charge Door, lb/hr/ton – 1693
CH_4 into Afterburner, lb/hr/ton – 25
Total Gas to Gas Cleaning System, lb/hr/ton – 6167

While the major function of the afterburner is the oxidation of carbon monoxide for safety and environmental control, two other processes may also occur. If oily scrap is charged to the cupola, the cupola exit gases will contain oil aerosol. This oil will be oxidized in the afterburner, to the benefit of the particulate collection devices which follow. There is also combustible dust in the exhaust gases, consisting mostly of coke fines (carbon). This, too, is burned to some extent in the afterburner.

The gases leaving the afterburner are too hot to go directly into a particulate collector without cooling. Three types of coolers are conventionally used: quenchers, evaporative coolers, and radiant coolers. Quenchers are distinguished from evaporative coolers in that quenchers are devices that adiabatically saturate using an excess amount of water, whereas evaporative coolers control the amount of cooling by controlling the amount of water available for evaporation. Cooling by air dilution is not widely practiced because of the relatively large volumes required and the attendant increase in the size of the collection device. There is a current EPA demonstration project on recuperative heat exchange for an iron foundry cupola,(5) however, the method is not used conventionally.

The choice between radiant and direct contact water cooling depends upon several variables:

1. *The nature of the abatement equipment downstream.* If a fabric collector is used, cooling in a quencher could lead to wetting and blinding the bags.

2. *The temperature of the exhaust gas leaving the cupola.* Radiant coolers are efficient only at high temperature levels. Inlet temperatures which are too high, however, can lead to metallurgy problems.

3. *Costs of utilities and of maintenance labor.* Quenchers and evaporative coolers consume water, may need more maintenance than radiant coolers, and add to the total gas volume to be handled.

4. *Capital cost.* Radiant coolers may cost many times more than quenchers or evaporative coolers.

5. *Space.* Radiant coolers require more space than quenchers or evaporative coolers.

For most applications, quenchers or evaporative coolers will be the more attractive choice because of the lower capital cost.

Collection of particulate matter emitted by cupolas can be accomplished efficiently by fabric filters, wet scrubbers, and electrostatic precipitators. Precipitators are usually uneconomical, however, because the exhaust gas rates are too low.

The design of the collector will be influenced by the size distribution of the particles which are emitted. The distribution varies greatly from cupola to cupola. Data which demonstrate the extent of this variation are shown in Table 216.

Fabric filters are effective devices in this application and have been used in many commercial installations. The inlet gas temperature to the fabric filter must be closely controlled, however, to prevent damage to the bags from high temperature upsets. Fabric filter systems should have a bypass which is activated by a high temperature alarm in the inlet gas stream. This bypass may be the cupola closure cap. Carry-over of incandescent particles or sparks must also be guarded against in the design of the fabric filter system.

Wet scrubber systems have also been successfully applied as control devices on cupolas. In most cases, they require a high energy input due to the small size of the particulate matter emitted from the cupola. In addition, they are subject to acid corrosion resulting from the dissolving of sulfur oxides from the cupola top gases in the scrubber recycle liquor. Where sulfur oxides exist in significant quantity, stainless steels such as 316L must be used in the scrubber to resist attack.

Table 217 presents a summary of abatement equipment type, and number of units of that particular type, that have been used for collecting particulate matter from foundries.

SPECIFICATIONS AND COSTS

Equipment specifications have been written for both fabric filters and wet scrubbers. In each case, the specification is based upon cupola exhaust gas leaving the afterburner. The system for both types of equipment includes a gas cooler followed by the collector and auxiliary equipment. Two different types of gas coolers were specified, one for each type of collection equipment. The gas cooling equipment appropriate to each collection device is a legitimate part of the cost of the abatement system, because no cooling of the gases is required where collection of particulate matter is not required.

Specifications for the fabric filter system are listed in Tables 218, 219 and 220. Vendors' quotations were not obtained for the two sizes specified in Table 220. The cost data for these two sizes were determined by interpolation and extrapolation of the cost data obtained for the two sizes specified in Table 219.

TABLE 216

PARTICLE SIZE DISTRIBUTIONS OF EMISSIONS FROM THREE DIFFERENT CUPOLAS

Wt.% Less Than	A	B	C
200 microns		99	
100 microns		99	
50 microns		92	60
20 microns	99	34	55
10 microns	98	12	45
5 microns	82	2	28
2 microns	64		13

TABLE 217

ABATEMENT EQUIPMENT TYPES USED FOR COLLECTING PARTICULATE MATTER FROM FOUNDRIES[(6)] *

Abatement Equipment Type	Number of Units Installed	Number of Foundries
Fabric Filter	5,216	2,198
Cyclones	1,216	514
Wet Scrubber	2,170	778
Electrostatic Precipitator	50	32
In Plant Dust Collection	2,629	1,122

*Data represents all foundries.

Both sets of specifications include an evaporative cooler operating at a maximum outlet temperature of 400° F. Cost data for the system are presented on Tables 221 and 222 and on Figures 172 and 173. Also presented on these tables and figures are the interpolated and extrapolated costs determined for the two fabric filter sizes specified in Table 220. Confidence limits for the capital cost data obtained for the two fabric filter sizes specified in Table 219 are presented on Table 223 and Figure 174.

Specifications for the wet scrubber system are shown on Tables 223 and 224. The specifications include a quencher type gas cooler followed by a venturi scrubber. For the high efficiency cases, the scrubber is followed by an aftercooler which reduces the horsepower requirement on the system fan by condensation. Cost data for the system are presented on Tables 225 and 226 and on Figures 175, 176, 177 and 178. Confidence limits for the capital costs of the high collection efficiency cases are presented on Table 228 and Figure 179.

Bids were not obtained for afterburners; however, generalized costs were obtained through private communications.(8) The burners range in price from $5,000 to $10,000/cupola.

The installation cost was estimated to be $2,100. The burner duty ranges from 5,200,000 Btu/hr/cupola to 16,000,000 Btu/hr/cupola. This corresponds to an annual fuel cost range of $21,900/yr/cupola to $67,000/yr/cupola. The fuel requirement is quite varied for cupola afterburners and in many cases the combustion of carbon monoxide could sustain itself.

Generalized cost breakdown for afterburners is as follows:

Device Cost Range	$ 5,000 to $10,000/cupola
Installation Cost	$ 2,100 to $2,100/cupola
Total Cost Range	$ 7,000 to $12,000/cupola
Fuel Cost Range	$21,900 to $67,500/yr/cupola

TABLE 218

FABRIC FILTER PROCESS DESCRIPTION FOR GRAY IRON FOUNDRY CUPOLA SPECIFICATION

Emissions from a pair of new gray iron cupolas are to be controlled with a fabric filter. The cupolas are to operate alternately, so the emissions from only one need be treated. The cupolas are conventional refractory lined units. Gas takeoffs are above the charge doors. Products of combustion and charge door indraft are preheated to a degree sufficient to insure that the gases leaving the cupola and entering the pollution control system will be 1,200°F. Each cupola is operated for 24 hours, then shut down for maintenance while the other cupola is operating. The operating factor for the abatement system serving the cupolas is 5,280 hr/yr.

The abatement system will consist of:

1. *An evaporative cooler*
2. *A fabric filter*

An afterburner will be supplied by others. It will be located in the stack above the charge door. The geometry of the installation will be arranged to insure sufficient turbulence for complete combustion. Stack inserts to achieve turbulence may be used if necessary.

The evaporative cooler will be used to lower the temperature of the afterburner exhaust gas to the required inlet temperature of the fabric collector. The cooler will be designed for a maximum inlet temperature of 2,200°F. Sufficient space is available for the installation of the evaporative cooler at a distance of 15 feet from the cupola. The fabric filter will be located 40 feet from the evaporative cooler. The afterburner exhaust gases leave the top of the cupola stack at a height of 50 feet above grade.

The fabric filter will be located immediately downstream of the evaporative cooler. The inlet gas stream will be monitored with a high temperature alarm and will be diverted out the cupola stack in the event of an alarm condition. The filter will be either a shaker or reverse collapse type unit designed with a minimum of four isolatable compartments. The unit will operate under either positive or negative pressure and will be equipped with screw conveyors and air locks for solids removal. A dust storage bin with a capacity of two days operation will be provided at the discharge of the screw conveyors. The system fan will be sized with at least 20% excess capacity when operating at the design pressure drop of the entire abatement system and 90% of the maximum recommended speed.

TABLE 219

FABRIC FILTER OPERATING CONDITIONS FOR GRAY IRON FOUNDRY CUPOLA SPECIFICATION

Two sizes of fabric filters are specified at one efficiency level. Vendors' quotations should consist of one quotation for each of the two sizes, with a representation of the efficiency expected for the unit quoted. The efficiency quoted may be better than the efficiency called for in this specification.

	Small	*Large*
Cupola Inside Diameter, in.	36	90
Cupola Melt Rate, ton/hr	6.6	39.5
Cupola Iron/Coke Ratio	10	10
Cupola Process Weight, lb/hr	14,730	88,560
Afterburner Exhaust Gas		
ACFM	27,500	145,000
Temperature, °F	2,000	2,000
SCFM	5,920	31,300
Solids Loading		
lb/hr	108	540
gr/ACF	0.459	0.434
gr/DSCF	2.42	2.29
Cooler Outlet Gas		
ACFM	14,600	77,100
Temperature, °F	400	400
SCFM	9,000	47,500
Solids Size Distribution		
% > 10 microns	68	68
% < 10 microns	32	32
% < 5 microns	24	24
% < 1 micron	11	11
% < 0.7 micron	8	8
% < 0.3 micron	1	1
Fabric Filter Inlet Gas		
ACFM	14,600	77,100
Temperature, °F	400	400
SCFM	9,000	47,500
Case 1 – High Efficiency		
Solids Loading		
lb/hr	1.25	6.61
gr/ACF	0.01	0.01
gr/DSCF	0.018	0.018
Collector Efficiency, wt.%	98.8	99.2
Air-to-cloth ratio	2.0/1	2.0/1

TABLE 220

FABRIC FILTER OPERATING CONDITIONS FOR

GRAY IRON FOUNDRY CUPOLAS

Two sizes of fabric filters are specified. Vendors' quotations were not obtained for either of these sizes. Cost data for these two sizes were determined by extrapolating the cost data obtained for the two fabric filter sizes specified in Table 11.

	Small	*Large*
Cupola Inside Diameter, in.	48	114
Cupola Melt Rate, ton/hr	10	60
Cupola Iron/Coke Ratio	10	10
Cupola Process Weight, lb/hr	22,400	134,400
Afterburner Exhaust Gas		
ACFM	41,800	220,000
Temperature, °F	2,000	2,000
SCFM	9,000	47,500
Solids Loading		
lb/hr	108	540
gr/ACF	0.302	0.286
gr/DSCF	1.60	1.50
Cooler Outlet Gas		
ACFM	22,150	116,930
Temperature, °F	400	400
SCFM	13,650	72,060
Fabric Filter Inlet Gas		
ACFM	22,150	116,930
Temperature, °F	400	400
SCFM	13,650	72,060
Case 1 – High Efficiency		
Solids Loading		
lb/hr	1.9	10
gr/ACF	0.01	0.01
gr/DSCF	0.018	0.018
Collector Efficiency, wt.%	98.0	98.3
Air-to-cloth ratio	2/1	2/1

Preceding page blank

TABLE 221

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR FABRIC FILTERS FOR GRAY IRON FOUNDRY CUPOLAS

	Extrapolated		Specified	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	41,800	220,000	27,500	145,000
°F	2,000	2,000	2,000	2,000
SCFM	9,000	47,500	5,920	31,300
Moisture Content, Vol. %	2	2	2	2
Effluent Contaminant Loading				
gr/ACF (ppm)	0.302	0.286	0.459	0.434
lb/hr	108	540	108	540
Cleaned Gas Flow				
ACFM	22,150	116,930	14,600	77,100
°F	400	400	400	400
SCFM	13,650	72,060	9,000	47,500
Moisture Content, Vol. %	35.4	35.4	35.4	35.4
Cleaned Gas Contaminant Loading				
gr/ACF (ppm)	0.01	0.01	0.01	0.01
lb/hr	1.9	10.05	1.25	6.61
Cleaning Efficiency, %	98.0	98.3	98.8	99.2
(1) Gas Cleaning Device Cost	37,000*	140,000*	26,600	99,850
(2) Auxiliaries Cost	45,000*	170,000*		
(a) Fan(s)			4,500	21,600
(b) Pump(s)			375	2,375
(c) Damper(s)			1,550	11,925
(d) Conditioning, Equipment			21,000	65,500
(e) Dust Disposal Equipment			2,950	21,750
(3) Installation Cost	103,000*	390,000*		
(a) Engineering			3,700	11,700
(b) Foundations & Support Ductwork Stack Electrical Piping Insulation Painting Supervision Startup Performance Test Other			67,230	218,125
(4) Total Cost	185,000*	700,000*	127,905	452,825

*Extrapolated from Cost Data presented in Figure 6

FIGURE 172

CAPITAL COSTS FOR FABRIC FILTERS FOR GRAY IRON FOUNDRY CUPOLAS

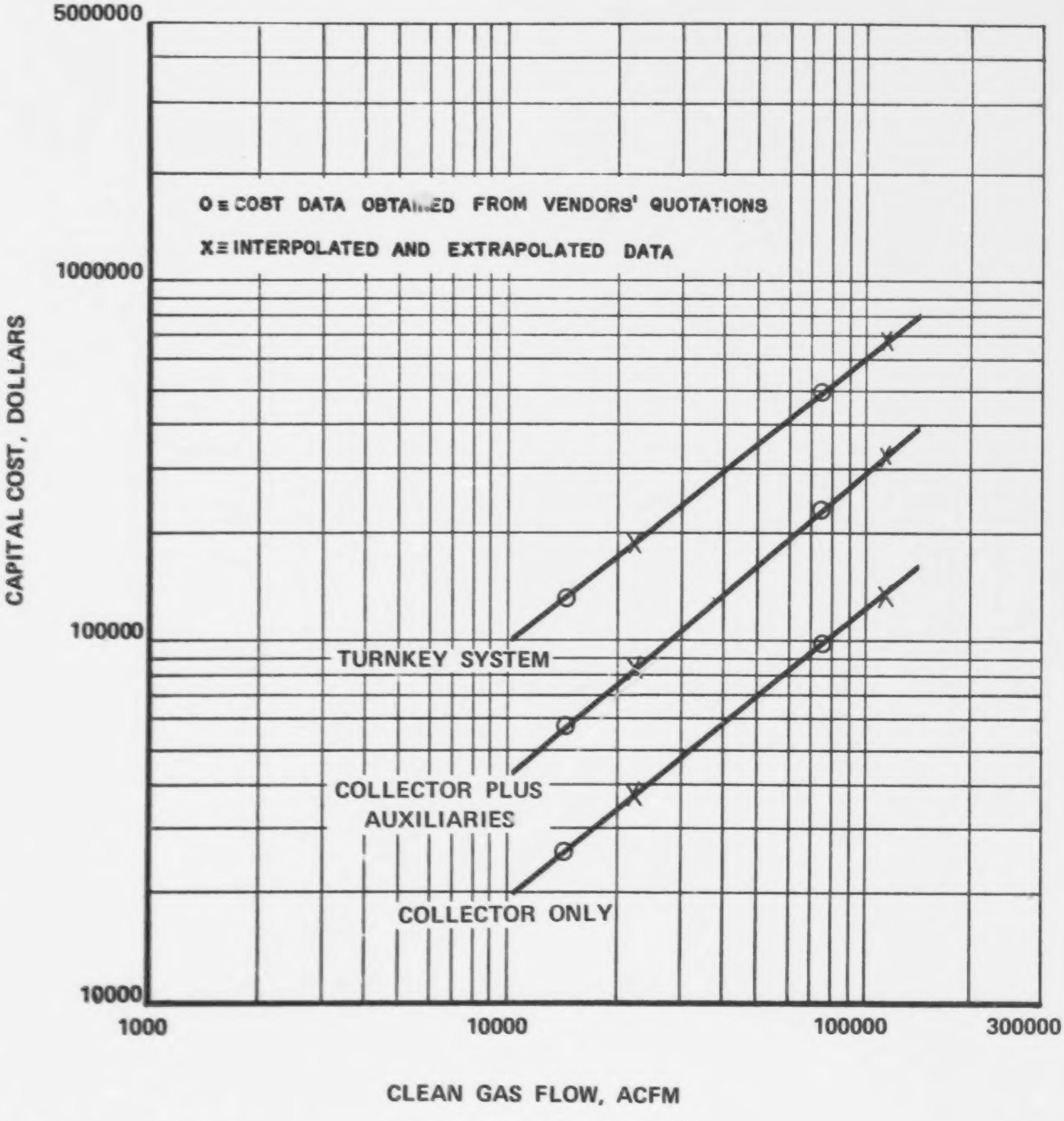


TABLE 222

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR FABRIC FILTERS FOR GRAY IRON FOUNDRY CUPOLAS

Operating Cost Item	Unit Cost	Extrapolated		Specified	
		22,150 ACFM	116,930 ACFM		
Operating Factor, Hr/Year		5,280	5,280	5,280	5,280
Operating Labor (if any)					
Operator	$6/hr			3,950	3,950
Supervisor	$8/hr			1,050	1,050
Total Operating Labor				5,000	5,000
Maintenance					
Labor					
Materials					
Total Maintenance				4,980	4,980
Replacement Parts					
Total Replacement Parts				6,000	16,500
Utilities					
Electric Power	$.011/kw-hr			2,850	16,900
Fuel					
Water (Process)	$.25/M gal			2,400	6,750
Water (Cooling)					
Chemicals, Specify					
Total Utilities				5,250	23,650
Total Direct Cost		26,000*	60,000*	21,230	50,130
Annualized Capital Charges		18,500	70,000	12,790	45,282
Total Annual Cost		44,500	130,000	34,020	95,412

*Extrapolated from Operating Cost Data presented in Figure 7

FIGURE 173

ANNUAL COSTS FOR FABRIC FILTERS FOR GRAY IRON FOUNDRY CUPOLAS

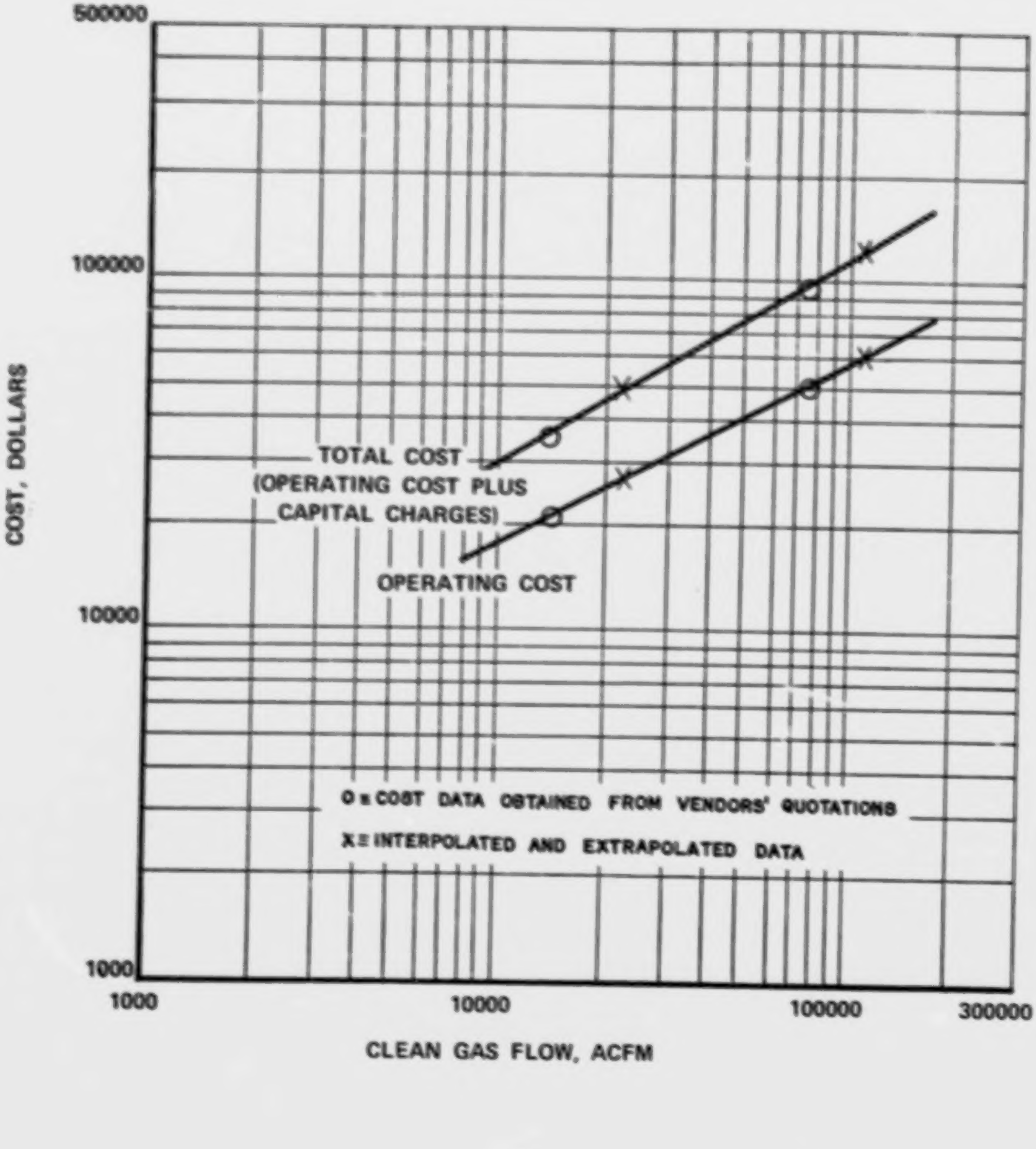


TABLE 223

CONFIDENCE LIMITS FOR CAPITAL COST OF FABRIC FILTERS FOR GRAY IRON FOUNDRY CUPOLAS

Population Size – 20 Sample Size – 2

Capital Cost = $127,905

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
90	$124,829	$130,981
75	121,092	134,718

Capital Cost = $452,825

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
90	$350,177	$555,473
75	225,503	680,147

FIGURE 174

CONFIDENCE LIMITS FOR CAPITAL COST OF FABRIC FILTERS FOR GRAY IRON FOUNDRY CUPOLAS

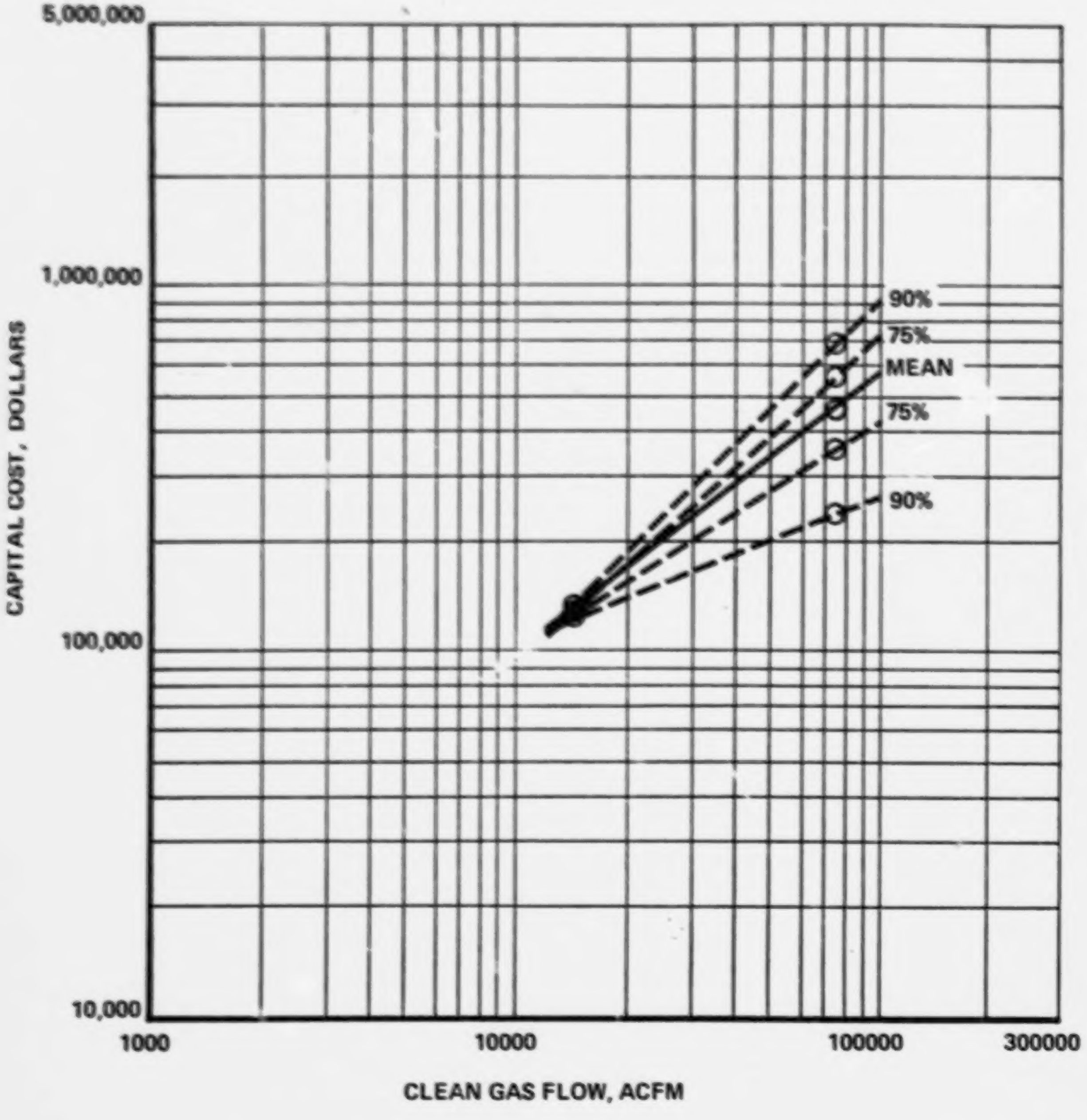


TABLE 224

WET SCRUBBER PROCESS DESCRIPTION FOR GRAY IRON CUPOLA SPECIFICATION

Emissions from a pair of new gray iron cupolas are to be controlled with a wet scrubber. The cupolas are to operate alternately, so the emissions from only one need be treated. The cupolas are conventional refractory lined units. Gas takeoffs are above the charge doors. Products of combustion and charge door indraft are preheated to a temperature sufficient that the gases leaving the cupola and entering the pollution control system will be 1,200°F. Each cupola is operated for 24 hours then shutdown for maintenance while the other cupola is operating. The yearly operating factor for the pollution control system is 5,280 hr/yr.

The abatement system will consist of:

1. A quencher
2. A wet scrubber

An afterburner will be supplied by others. It will be located in the stack above the charge door. The geometry of the installation will be arranged to insure sufficient turbulence for complete combustion. Stack inserts to achieve turbulence may be used if necessary.

The quencher will be used to lower the temperature of the afterburner exhaust gas to the inlet temperature of the wet scrubber. The quencher should be located immediately after the gases leave the top of the cupola. The quencher should be designed to lower the gas temperature from a maximum of 2,200°F to less than 200°F scrubber inlet temperature. The cupola gas leaves the unit at a height of 50 feet above grade. The duct after the quencher must transport the cooled gas to the scrubber inlet located 40 feet away from the quencher. The quencher is 15 feet away from the cupola.

The wet scrubber will be a venturi-type with a liquid-to-gas ratio in excess of 5 gpm/1,000 ACFM (saturated). Materials of construction will be 316L stainless steel or equivalent. The system will include the following:

1. A fan, located downstream of the scrubber sized with 20% excess capacity when operating at the design pressure and 90% of the maximum recommended speed.
2. A slurry settling assembly capable of producing a thickened underflow (30% solids) while returning water with minimum solids content.
3. A minimum of at least two filters to dewater the slurry product which are capable of producing a filter cake with ≥ 65% solids content suitable for open truck transportation.
4. An aftercooler capable of lowering the scrubber effluent gas temperature to 105°F by countercurrent contact with 90°F cooling water (Case II – High Efficiency only).

TABLE 225

WET SCRUBBER OPERATING CONDITIONS FOR GRAY IRON CUPOLA SPECIFICATION

	Small	*Large*
Cupola Inside Diameter, in.	*48*	*114*
Cupola Melt Rate, ton/hr	*10*	*60*
Cupola Iron/Coke Ratio	*10*	*10*
Cupola Process Weight, lb/hr	*22,400*	*134,400*
Afterburner Exhaust Gas		
ACFM	*41,800*	*220,000*
Temperature, °F	*2,000*	*2,000*
SCFM	*9,000*	*47,500*
DSCFM	*8,820*	*46,550*
Water Content, vol%	*2*	*2*
Solids Loading		
lb/hr	*108*	*540*
gr/ACF	*0.301*	*0.286*
gr/DSCF	*1.429*	*1.353*
Quencher Outlet Gas		
ACFM	*18,700*	*98,300*
Temperature, °F	*200*	*200*
Water Content, vol%	*41*	*41*
SCFM	*15,000*	*73,900*
DSCFM	*8,820*	*46,550*
Solids Loading		
lb/hr	*108*	*540*
gr/ACF	*0.21*	*0.20*
gr/DSCF	*1.429*	*1.353*
Solids Size Distribution		
% > 10 microns	*68*	*68*
% < 10 microns	*32*	*32*
% < 5 microns	*24*	*24*
% < 1 micron	*11*	*11*
% < 0.7 micron	*8*	*8*
% < 0.3 micron	*1*	*1*
Scrubber Outlet Gas		
ACFM	*18,100*	*95,500*
Temperature, °F	*171*	*171*
Water Content, vol %	*42*	*42*
SCFM	*15,200*	*80,200*
DSCFM	*8,820*	*46,550*
Aftercooler Outlet Gas		
ACFM	*10,300*	*54,400*
Temperature, °F	*105*	*105*
SCFM	*9,660*	*51,000*
Water Content, vol %	*8.7*	*8.7*
DSCFM	*8,820*	*46,550*

	Small	Large
Case I – Medium Efficiency		
Solids Loading		
lb/hr	20.4	40
gr/ACF	0.16	0.06
gr/DSCF	0.27	0.10
Collection Efficiency, wt.%	81.1	92.6
Scrubber, Δ P	25 in. w.c.	25 in. w.c.
Case II – High Efficiency		
Solids Loading		
lb/hr	0.9	4.7
gr/ACF	0.010	0.010
gr/DSCF	0.011	0.011
Collection Efficiency, wt.%	99.1	99.1
Scrubber, Δ P	60 in. w.c.	60 in. w.c.

Preceding page blank

TABLE 226

ESTIMATED CAPITAL COST DATA (COSTS IN DOLLARS) FOR WET SCRUBBERS FOR GRAY IRON FOUNDRY CUPOLAS

	LA Process Wt.		High Efficiency	
	Small	Large	Small	Large
Effluent Gas Flow				
ACFM	41,800	220,000	41,800	220,000
°F	2,000	2,000	2,000	2,000
SCFM	9,000	47,500	9,000	47,500
Moisture Content, Vol. %	2	2	2	2
Effluent Contaminant Loading				
gr/ACF	0.301	0.286	0.301	0.286
lb/hr				
Cleaned Gas Flow				
ACFM	18,100	95,500	10,300	54,400
°F	171	171	105	105
DSCFM	8,820	46,550	8,820	46,550
Moisture Content, Vol. %	42	42	8.7	8.7
Cleaned Gas Contaminant Loading				
gr/ACF	0.16	0.06	0.01	0.01
lb/hr	20.4	40.0	0.9	4.7
Cleaning Efficiency, %	81.1	92.6	99.1	99.1
(1) Gas Cleaning Device Cost	19,067	55,833	22,567	69,000
(2) Auxiliaries Cost (a) Fan(s) (b) Pump(s) (c) Damper(s) (d) Conditioning, Equipment (e) Dust Disposal Equipment	56,700	110,120	71,833	149,567
(3) Installation Cost (a) Engineering (b) Foundations & Support Ductwork Stack Electrical Piping Insulation Painting Supervision Startup Performance Test Other	117,166	203,467	131,333	239,800
(4) Total Cost	192,933	369,420	225,733	458,367

TABLE 227

ANNUAL OPERATING COST DATA (COSTS IN $/YEAR) FOR WET SCRUBBERS FOR GRAY IRON FOUNDRY CUPOLAS

Operating Cost Item	Unit Cost	LA Process Wt.		High Efficiency	
		Small	Large	Small	Large
Operating Factor, Hr/Year		5,280	5,280	5,280	5,280
Operating Labor (if any)					
Operator	$6/hr	4,560	4,560	4,560	4,560
Supervisor					
Total Operating Labor		4,560	4,560	4,560	4,560
Maintenance					
Labor					
Materials					
Total Maintenance		5,923	10,248	6,898	12,398
Replacement Parts					
Total Replacement Parts		4,290	9,300	5,415	11,770
Utilities					
Electric Power	$.011/kw-hr	7,969	63,629	17,389	121,055
Fuel					
Water (Process)	$.25/M gal	1,520	9,738	2,080	9,738
Water (Cooling)	$.05/M gal			6,490	37,734
Chemicals, Specify					
Total Utilities		9,489	73,367	25,959	168,527
Total Direct Cost		24,262	97,475	42,832	197,255
Annualized Capital Charges		19,293	36,942	22,573	45,837
Total Annual Cost		43,555	134,417	65,405	243,092

FIGURE 175

CAPITAL COSTS FOR WET SCRUBBERS FOR GRAY IRON FOUNDRY CUPOLAS (MEDIUM EFFICIENCY)

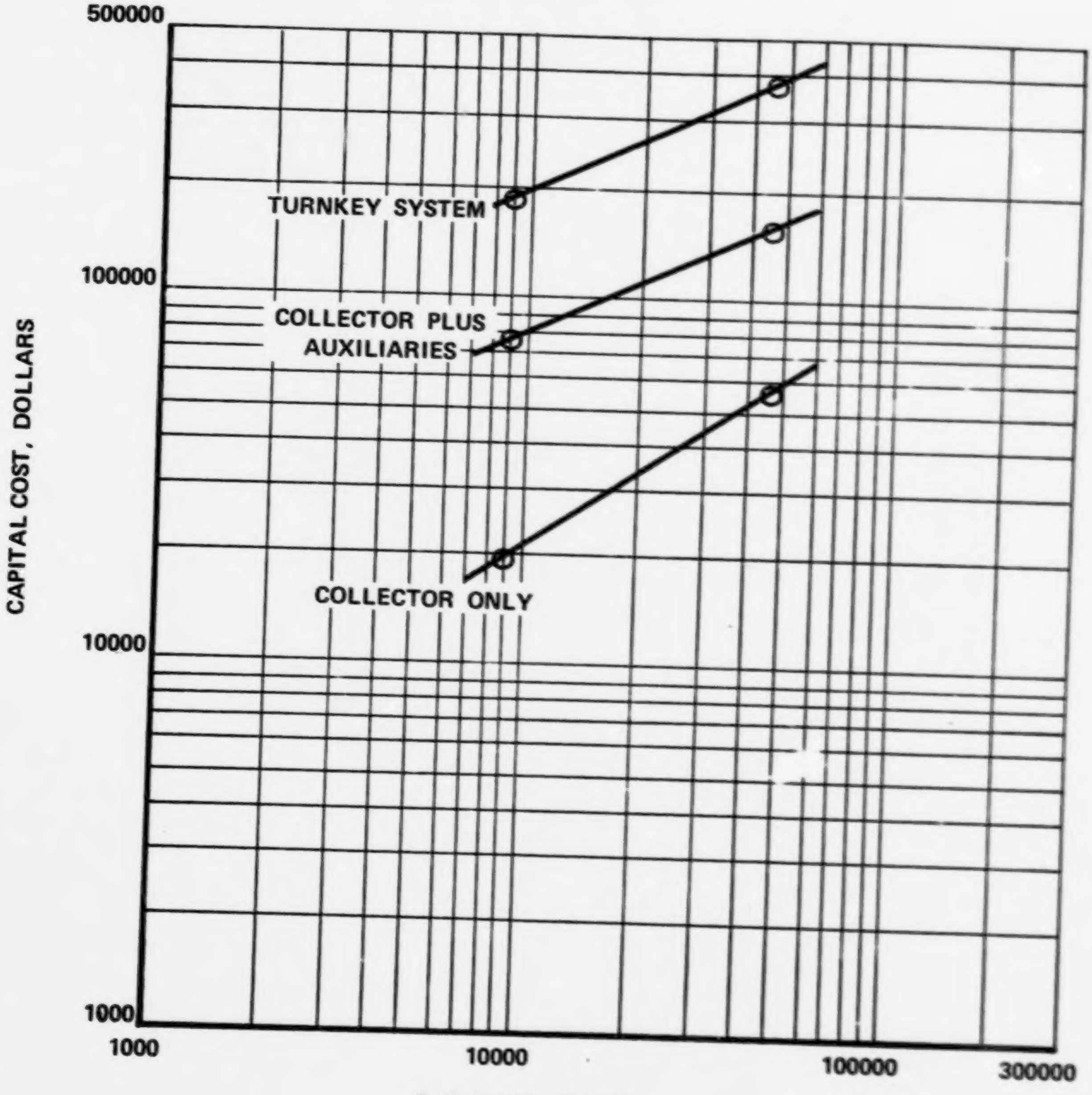


FIGURE 176

ANNUAL COSTS FOR WET SCRUBBERS
FOR GRAY IRON FOUNDRY CUPOLAS
(MEDIUM EFFICIENCY)

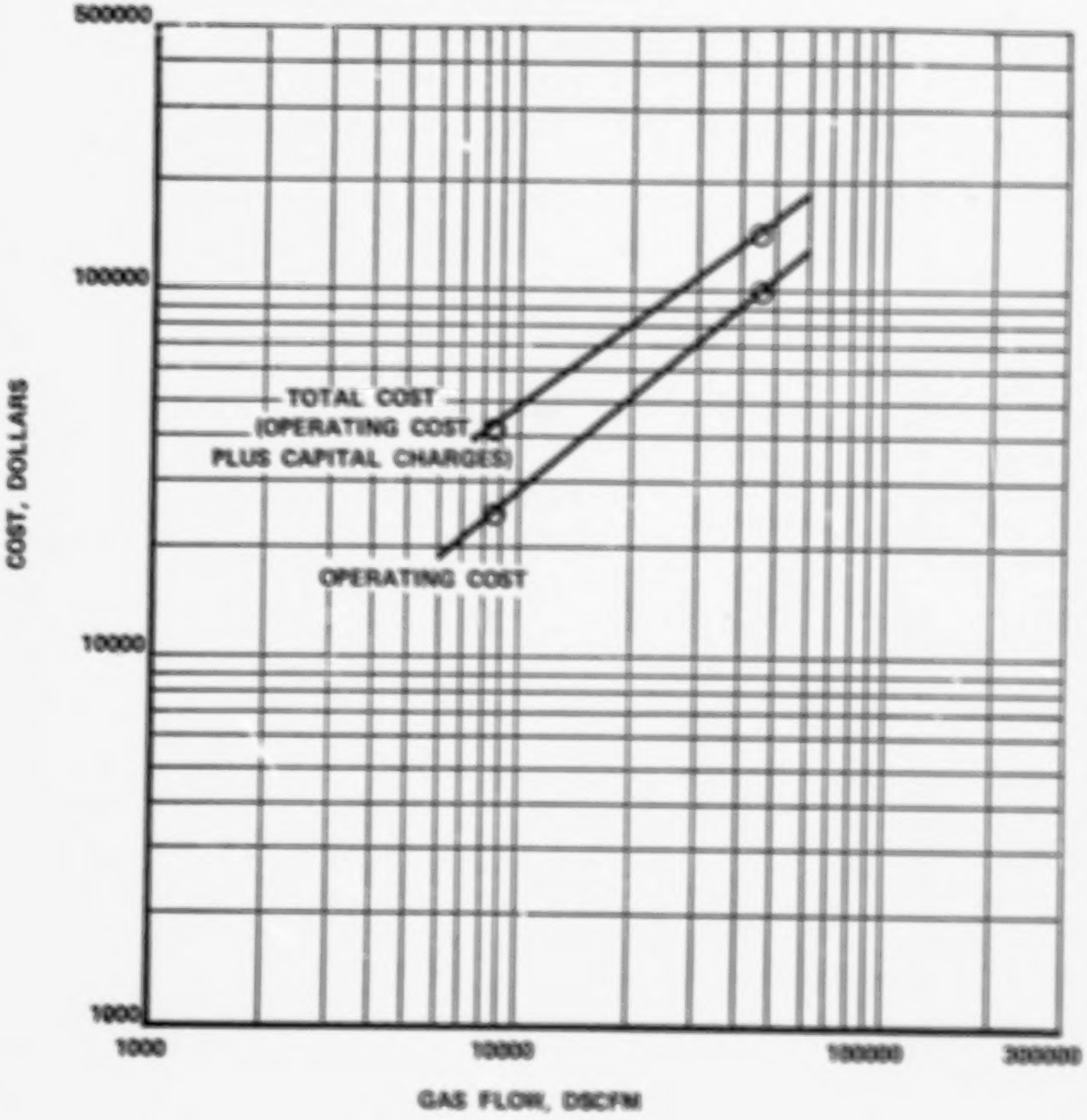


FIGURE 177

CAPITAL COSTS FOR WET SCRUBBERS
FOR GRAY IRON FOUNDRY CUPOLAS
(HIGH EFFICIENCY)

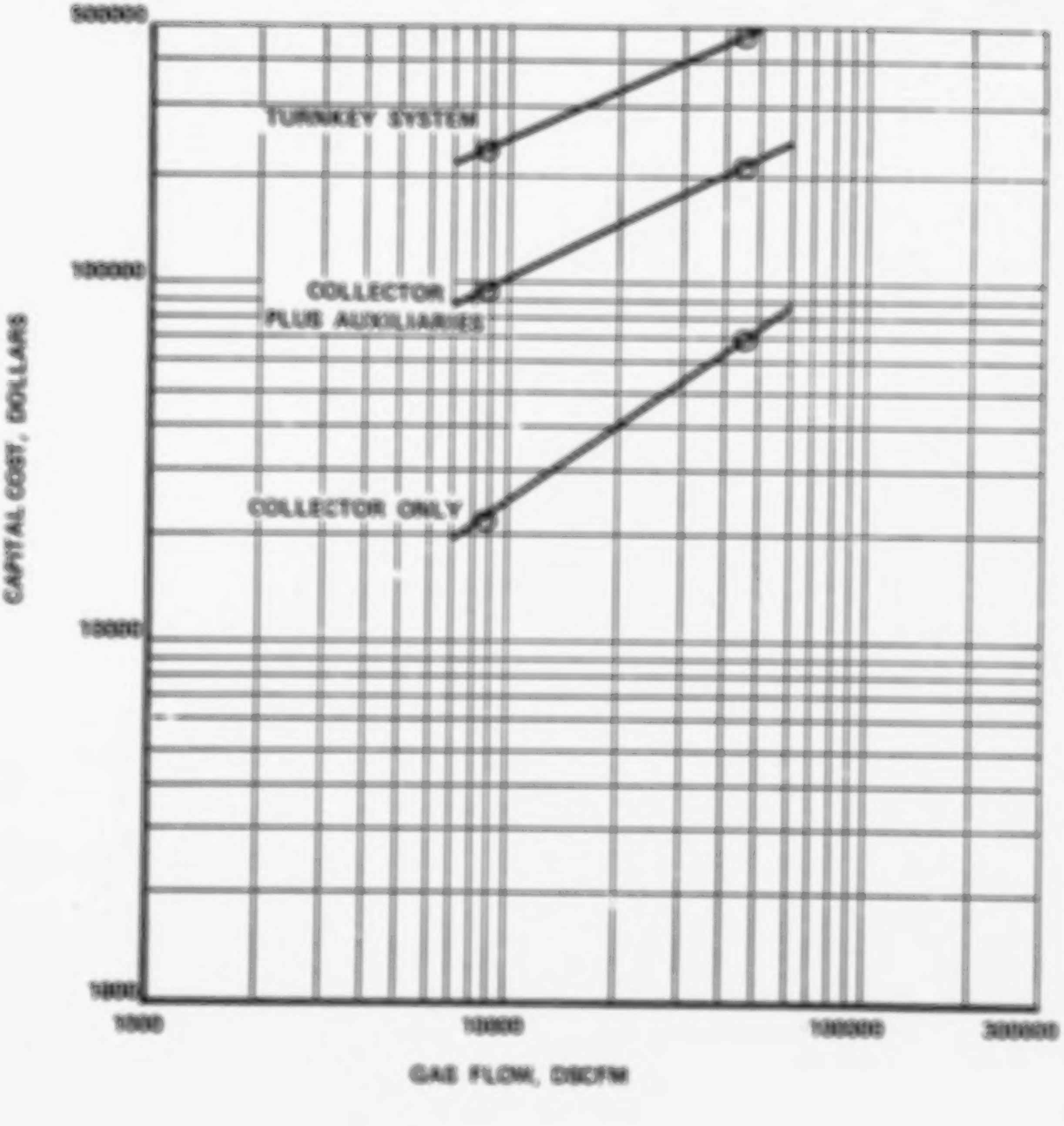


FIGURE 178

ANNUAL COSTS FOR WET SCRUBBERS
FOR GRAY IRON FOUNDRY CUPOLAS
(HIGH EFFICIENCY)

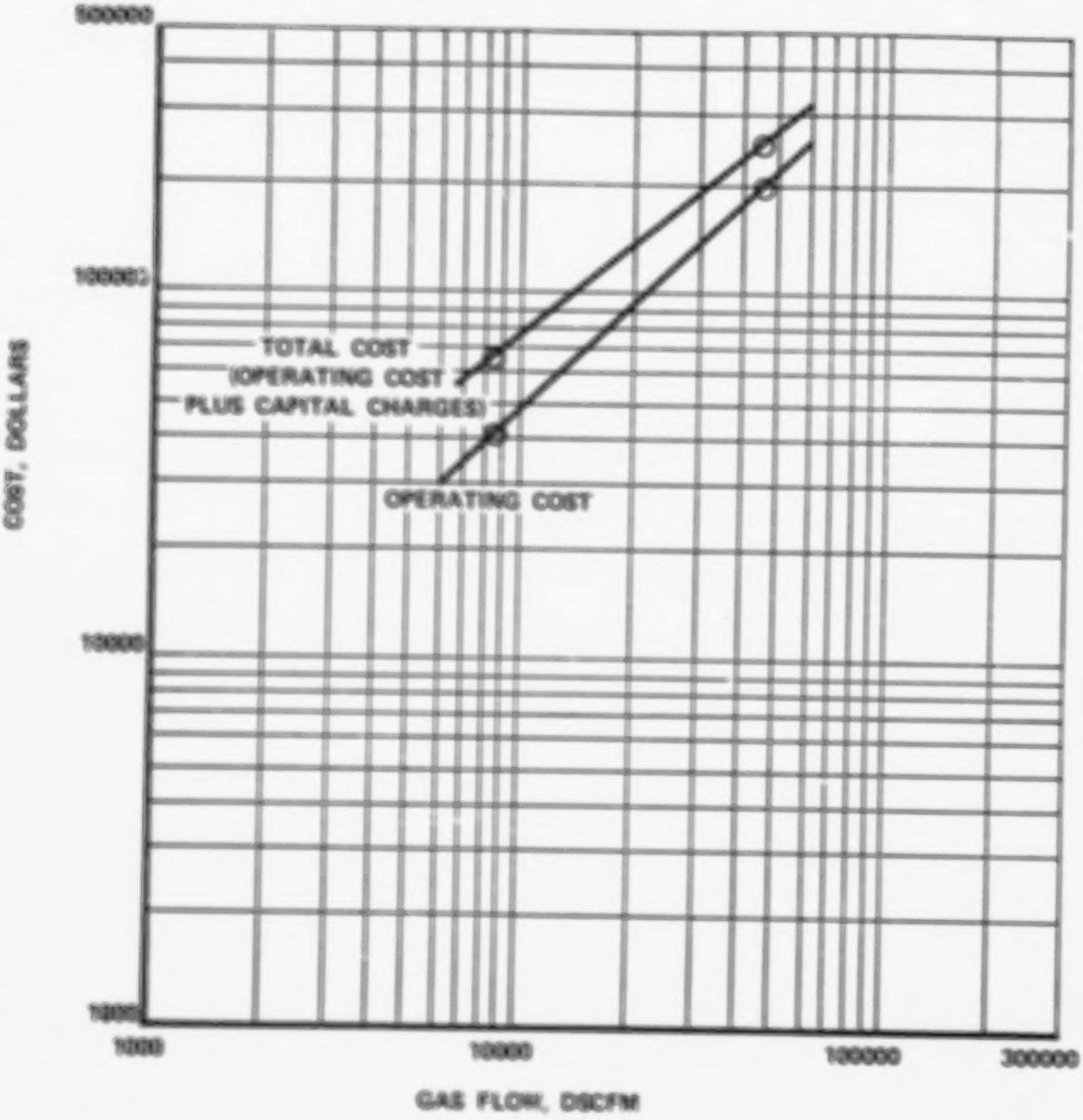


TABLE 228

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR GRAY IRON FOUNDRY CUPOLAS (HIGH EFFICIENCY)

Population Size – 20 Sample Size – 3

Capital Cost = $225,733

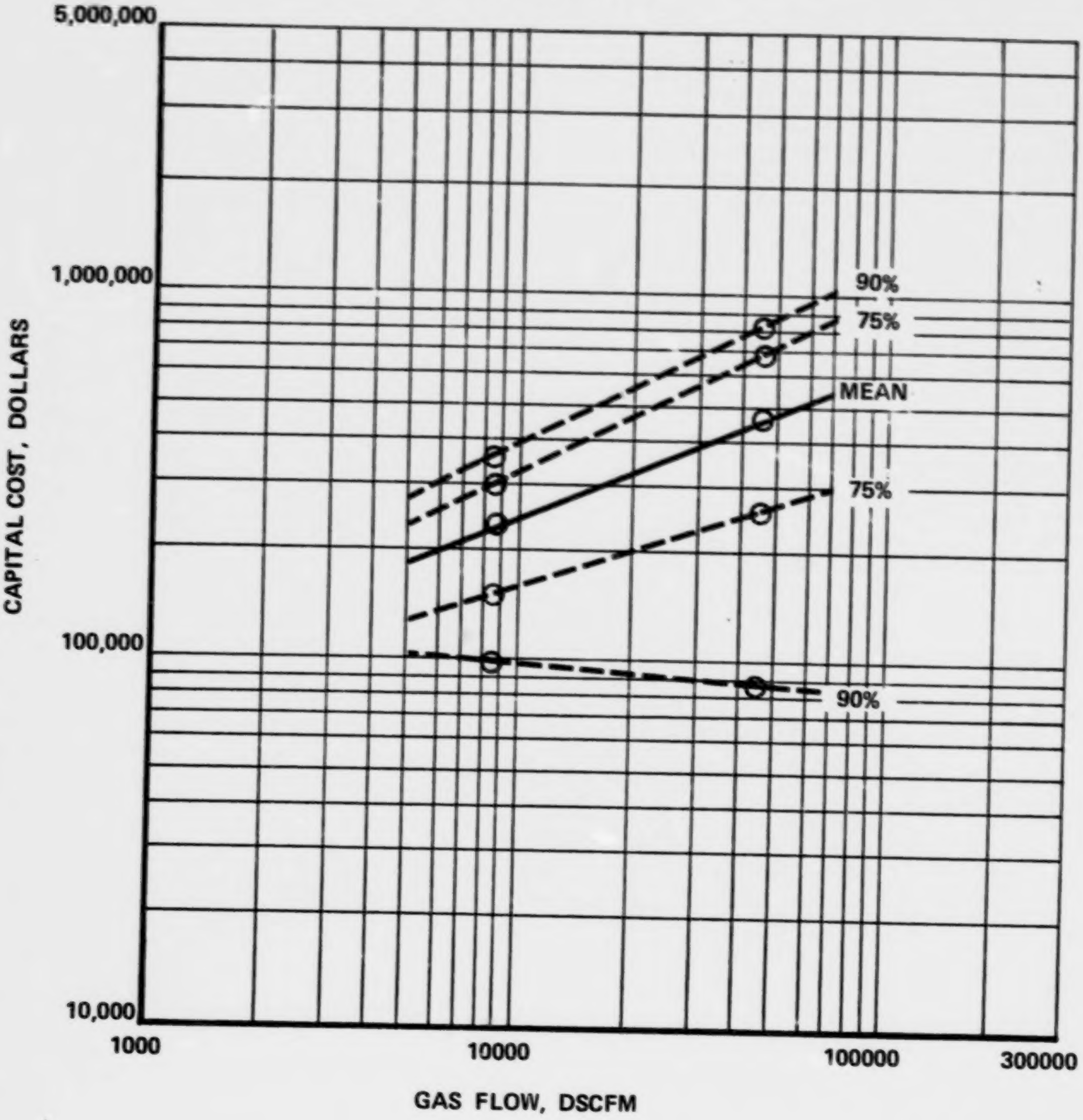
	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$187,854	$263,613
75	152,339	299,128
90	97,689	353,777

Capital Cost = $458,367

	Capital Cost, Dollars	
Conf. Level, %	Lower Limit	Upper Limit
50	$348,419	$568,315
75	245,333	671,401
90	86,709	830,024

FIGURE 179

CONFIDENCE LIMITS FOR CAPITAL COST OF WET SCRUBBERS FOR GRAY IRON FOUNDRY CUPOLAS (HIGH EFFICIENCY)



REFERENCES

1. Systems Analysis of Emissions and Emissions Control in the Iron Foundry Industry, Vols. I, II, and III, A. T. Kearny and Co., Chicago, Ill., 1971. EPA Contract No. CPA 22-69-106

2. A Localized Study of Gray Iron Foundries to Determine Business and Technical Commonalities Conducive to Reducing Abatement Costs, J. A. Commins & Associates, Inc., Fort Washington, Pa., 1972. EPA Contract No. 68-04-0043

3. Air Pollution Engineering Manual, U.S. Dept. of Health, Education, and Welfare, Public Health Services Publication No. 999-AP-40, Cincinnati, Ohio, 1967.

4. Hultgren, Ralph, Fundamentals of Physical Metallurgy, Prentice Hall, Inc., New York, N.Y., 1952.

5. Iron Foundry Cupola-Recuperator Demonstration, Conceptual Test Program Definition, Catalytic, Inc., Charlotte, N.C., 1972. EPA Contract No. 68-02-0241.

6. Industry Census Guide, Metal Casting Journal, 1972 edition.

7. The Cupola and Its Operation, American Foundrymen's Society, 3rd edition, 1965.

8. Whiting Corporation, Harvey, Illinois 60426.

C. ADDITIONAL COST DATA

The previous section of this report dealt with the cost of air pollution control systems for eight specific processing applications. This section deals with the generalized correlation of costs for each type of control system for all eight of the process applications discussed. The discussion is divided into four parts:

1. A discussion of the annual operating cost basis, including both the direct and capital charge portions of this cost.
2. A discussion of the effects of utility price levels on overall costs.
3. Derivation of capital cost indices for each specific process application.
4. Generalized graphical correlations of capital and operating costs for each type of control system.

1. Discussion of Cost Basis

As noted in the introduction to this report, the total annual cost for a particular control system is the sum of the direct annual operating cost and an annualized capital charge.

In the previous section of this report the annual direct operating costs for air pollution control systems in specific processing applications were calculated using estimates supplied by the equipment manufacturers. These estimates were prepared in terms of the quantity of each operating cost item required, rather than the cost. A standard price was applied to these estimates by the coordinating engineer in order to determine the equivalent cost. The standard prices used are listed below:

Cost Item	Units	Price, $/Unit
Operating Labor		
Operator	man hrs	6
Supervisor	man hrs	8
Maintenance		
Materials	*	*
Labor	man hrs	6
Replacement Parts		
Utilities		
Electric Power	kwh	0.011
Fuel	mm Btu	0.80
Process Water	M Gal	0.25
Cooling Water	M Gal	0.05
Chemicals	*	

The sum of all the above items applied over a year's operation is the direct annual operating cost of the system.

The total annual cost of each system is calculated by adding the annualized capital charges to the direct annual operating cost. In calculating the annualized capital charges, the investment cost of the system, including taxes and interest, must be spread out over the useful life of the equipment. Many methods for annualizing investment cost are used. These methods fall into three major categories:

1. Straight line method which applies the capital charges at a fixed rate over the useful life of the control system.

2. Accelerated methods which apply the capital charges at a declining rate over the useful life, on the theory that aging or loss in value of equipment occurs to a greater degree on new equipment than on old equipment.

3. Methods which relate capital charges to some measure of equipment usage. These methods are seldom applied to processing equipment. The most common example is mileage-based depreciation of automobiles.

Of the two methods applicable to processing equipment, the most commonly used is the straight line method. This is the method used for the data presented in this report. Reasons for its common use are:

1. It is easy to understand and calculate.

2. It is thought by many to be the best approximation of the rate of obsolescence of process equipment.

3. It makes alternate control systems comparable on an annualized cost basis since the capital charges based upon this method are constant from year to year.

Once the decision has been made to use this method, the only critical issue is what value to use for the useful life of the control system. The useful life of any control system is, in reality, a composite of the useful lives of its component parts. Some of those parts have relatively long lives, others relatively short lives. The value chosen for the economic evaluation of a control system depends upon: the nature of the primary control device, the differences in expected useful life of similar equipment from different manufacturers, the maintenance practices of the owning firm, the battery limits defined for the system, the number and kind of structures built, and the accounting practices of the owning firm, among others. For these reasons, the value chosen will vary from firm to firm even for similar systems.

Taxes may also play a part in the determination of useful life. Under normal circumstances, control systems are depreciated over their normal useful lives. They may, however, be depreciated for tax purposes at an accelerated rate. Under certain circumstances, defined by the Internal Revenue Service, all or part of the air pollution control equipment may be amortized over a sixty month period. In most cases, this period is much shorter than the normal useful life. Accelerated depreciation for tax purposes, especially the sixty month amortization, has the effect of decreasing effective operating cost by deferring tax payments into the future. The discounted value of the cash outflow caused by the operation of the pollution control system is thereby reduced.

The money market at the time of equipment purchase is another important variable in the determination of capital charges. The rate at which money is available varies widely from firm to firm, as well as with overall economic considerations. The cost of capital for financing by means other than borrowing also varies over a wide range from firm to firm. Variations in the cost of financing can be large enough to affect the choice between alternative control systems.

For the purpose of presenting the annual operating cost data in this report, it was decided to use the same fixed percentage of total installed cost as the capital charge for all of the applications studied. The rate chosen was 10%. It was based upon an estimated useful equipment life of 15 to 20 years, debt capital availability at 6 to 8%, and a correction for the tax incentives available to installers of pollution control hardware of 2 to 4%. Although the rate chosen is a good general estimate, it does not purport to be the correct rate for any specific situation. It is used only as a good estimate to assist the cost presentations in this report.

2. Discussion of Utility Price Levels

Evaluation of, and selection among, equivalent control systems should be based upon both the capital cost and the operating cost. Part 1 of this section discussed the capital portion of the total annual operating cost and showed that the direct operating cost is composed of the following items:

Operating (operator and supervisor) Labor

Maintenance Labor and Materials

Replacement Parts

Utilities and Supplies

The utilities portion of the operating cost is a function of utility price levels. Price levels vary due to:

1. Geography – The price of natural gas, for example, is much higher in the New England states than in the Gulf Coast states.

2. Demand – The demand for low sulfur black fuel oil keeps its price as much as 50 cents per barrel higher than the equivalent higher sulfur fuel.

3. Nature of Use – The rates for interruptible gas service are lower than those for continuous service, and rates for peak period use of electrical service are as much as twice the rates for off-hour power consumption.

The total operating cost is the sum of the direct operating cost and the capital charges and is, therefore, a function of utility costs in varying degrees. The price level effect on the operating cost is proportional to the portion of the total cost which is due to utilities. These portions have been calculated for each type of control equipment considered in this report, and the results are presented in Table 229.

Table 229 shows that, on the average, utilities account for 44 to 73% of the total annual cost of wet scrubbers, while they account for only 13 to 18% of the total annual cost of electrostatic precipitators. Therefore, the price level of electrical power would effect the total annual operating cost of a wet scrubber more than a precipitator.

The effect of utility cost levels also varies within a given type of system. Capital cost is a larger part of the total cost of a small unit than it is of a large unit of the same type. Therefore, the utility costs are smaller relative to the total annual operating cost, and have less effect on it. Also, the operating parameters of a system affect the relative size of the utility costs. Scrubbers using a high pressure drop, for example, require more power to push the gas

TABLE 229

UTILITY COSTS

System	As Percent Of Direct Operating Cost			As Percent Of Total Annual Cost		
	Min.	Ave.	Max.	Min.	Ave.	Max.
Fabric Filters						
Small	18	42	64	11	23	41
Large	34	56	68	20	32	48
Incinerators w/o H. E.						
Small	53	79	95	33	59	84
Large	76	89	98	61	77	92
Incinerators with H. E.						
Small	33	65	96	16	47	83
Large	39	74	98	21	57	88
Wet Scrubbers						
Small	24	61	99.7	12	44	99.1
Large	28	73	99.9	28	73	99.9
Elec. Precipitators						
Small	26	38	48	10	13	17
Large	36	49	58	12	18	22
Carbon Adsorption						
Small	–	90	–	–	66	–
Large	–	93	–	–	72	–
Overall Average		66			48	

through, than do scrubbers having a low pressure drop. Therefore, utilities will be a larger portion of the total cost of high energy scrubbers than of low energy scrubbers.

Table 229 shows that for all the systems considered in this report the cost of utilities accounts, on the average, for about two thirds of the direct cost of operating a system, and for about one half of the total annual operating cost of a system. The general level of operating costs is substantial, and requires that price levels be considered when alternatives are being compared.

3. Derived Capital Cost Indices

For each of the process applications discussed in the previous sections of this report, capital costs of pollution control equipment have been presented for two or three different process sizes. This permits development of a mathematical expression for the capital cost of air pollution control systems as a function of process size in each application. The mathematical model chosen was the exponential form often used for relating cost and size of capital equipment.

$$\text{Capital Cost} = K\,(\text{Size})^{x}$$

Where

K and x are constants, and

Size is the capacity of the plant to which the abatement equipment is being applied.

This relationship assumes that a log-log plot of cost and size is a straight line for each application. For most types of equipment, this is a good assumption.

The constants K and x were evaluated by computer for each abatement application studied. Calculations were made for each of the three capital cost categories presented in each application:

1. Collector only
2. Collector plus auxiliaries
3. Turnkey system

Calculations were made using the computer program listed in Dartmouth Basic Language in Table 230.

The units of the "Size" term in the equation for each application are the same as those used in the prior discussion of that application. They are summarized in Table 231 .

The results of these calculations for generating capital costs in dollars, are presented in the following tables:

Process Area	Table Numbers
Phosphates	
– WPPA	232
– SPA	233
– GTSP	234
– DAP	235, 236, 237, 238
Feed and Grain	239, 240, 241, 242
Paint and Varnish	243, 244, 245, 246
Graphic Arts	247, 248, 249, 250, 251
Gray Iron Foundries	252, 253
Lime Kilns	254, 255, 256, 257, 258
Soap and Detergent	259, 260, 261

Also shown in these tables are the ratios of turnkey system cost to collector cost, total equipment cost to collector cost, and turnkey system cost to total equipment cost.

Calculated values of the exponents for the power function can be summarized by equipment type as follows:

(Text continued on page 644)

TABLE 230

```
120 FILES *
122 PRINT "FORM CONTROL"
124 INPUT Z
130 READ #1,N$
140 READ #1,E$,C$
145 IF END #1 THEN 650
150 IF J=1 GO TO 230
160 PRINT
170 PRINT
180 PRINT "                        DERIVED COST INDICES FOR ";N$
185 PRINT
187 PRINT
188 PRINT
190 PRINT
200 PRINT "           COLLECTOR TYPE";"            K*";"          X*";
201 PRINT "      B/A";"       C/A";"      C/B"
230 PRINT
240 FOR M = 1 TO 4
250 FOR N = 1 TO 2
260 READ #1, A(M,N)
270 NEXT N
280 NEXT M
290 FOR N = 1 TO 2
310 NEXT N
320 FOR M = 1 TO 3
330 X(M) = (LOG(A(M,1))-LOG(A(M,2)))/(LOG(A(4,1))-LOG(A(4,2)))
340 P(M) = (LOG(A(M,1))+LOG(A(M,2)))-X(M)*(LOG(A(4,1))+LOG(A(4,2)))
350 P(M) = EXP(P(M)/2)
360 NEXT M
370 FOR N = 1 TO 2
375 R(1,N)=A(2,N)/A(1,N)
380 R(2,N) = A(3,N)/A(1,N)
390 R(3,N)= A(3,N)/A(2,N)
400 NEXT N
410 PRINT "        ";E$
415 PRINT "        ";C$
420 PRINT,530,P(1),X(1)
430 PRINT,540,P(2),X(2)
440 PRINT,550,P(3),X(3)
450 PRINT
460 PRINT,560,R(1,1),R(2,1),R(3,1)
470 PRINT,570,R(1,2),R(2,2),R(3,2)
480 PRINT
500 J = 1
510 GOTO 140
530 FMT X10,"COLLECTOR ONLY(A)",X6,I7,X4,F6.3,X6,"-",X8,"-",X8,"-"
540 FMT X10,"TOTAL EQUIPMENT(B)",X5,I7,X4,F6.3,X6,"-",X8,"-",X8,"-"
550 FMT X10,"TURNKEY(C)",X13,I7,X4,F6.3,X6,"-",X8,"-",X8,"-"
560 FMT X12,"SMALL",X20,"-",X9,"-",X6,F6.3,X2,F7.3,X2,F6.3
570 FMT X12,"LARGE",X20,"-",X9,"-",X6,F6.3,X2,F7.3,X2,F6.3
580 PRINT
590 PRINT
600 PRINT "        *FOR USE IN EQUATION  COST = K*(SIZE)*EXP(X)"
610 FOR N = 1 TO 30
620 PRINT
630 NEXT N
640 GO TO 700
650 FOR Y=1 TO 12
660 PRINT
670 NEXT Y
680 GOTO 600
700 END
```

TABLE 231

UNITS OF PLANT SIZE FOR EACH PROCESS AREA

Process Area	Plant Size Units
Phosphates	
— WPPA	Ton/day P_2O_5
— SPA	Ton/day Product
— GTSP	Ton/day TSP
— DAP	Ton/day DAP
Feed and Grain	Lb/hr Product
Paint and Varnish	Gals Total Reactor Capacity
Graphic Arts	SCFM Exhaust Rate
Gray Iron Foundries	Ton/hr Cupola Melt Rate
Lime Kilns	Ton/day Kiln Capacity
Soap and Detergent	Lb/hr Product Rate

TABLE 232

DERIVED COST INDICES FOR WPPA PROCESS PLANTS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
PACKED C-F SCRUBBERS ONE-THIRD GAS FLOW					
COLLECTOR ONLY(A)	1709	.248	-	-	-
TOTAL EQUIPMENT(B)	2390	.269	-	-	-
TURNKEY(C)	11156	.200	-	-	-
SMALL	-	-	1.521	4.607	3.030
LARGE	-	-	1.539	4.477	2.908
PACKED C-F SCRUBBERS NORMAL GAS FLOW					
COLLECTOR ONLY(A)	2186	.337	-	-	-
TOTAL EQUIPMENT(B)	4421	.287	-	-	-
TURNKEY(C)	14195	.239	-	-	-
SMALL	-	-	1.402	3.331	2.382
LARGE	-	-	1.439	3.333	2.316

FOR USE IN EQUATION COST = K(SIZE)**EXP(X)

TABLE 233

DERIVED COST INDICES FOR SPA PROCESS PLANTS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
W-I VENTURI SCRUBBERS					
NOMINAL EFFICIENCY					
COLLECTOR ONLY(A)	738	.431	-	-	-
TOTAL EQUIPMENT(B)	738	.431	-	-	-
TURNKEY(C)	1460	.367	-	-	-
SMALL	-	-	1.000	1.439	1.439
LARGE	-	-	1.000	1.377	1.377
VENTURI-PACKED SCRUBBER					
NOMINAL EFFICIENCY					
COLLECTOR ONLY(A)	1405	.299	-	-	-
TOTAL EQUIPMENT(B)	3011	.249	-	-	-
TURNKEY(C)	6922	.249	-	-	-
SMALL	-	-	1.662	3.823	2.300
LARGE	-	-	1.605	3.692	2.300

FOR USE IN EQUATION: COST = K(SIZE)**EXP(X)

TABLE 234

DERIVED COST INDICES FOR GTSP PROCESS PLANTS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
VENTURI-PACKED SCRUBBER MOD. EFFICIENCY					
COLLECTOR ONLY(A)	322	.760	-	-	-
TOTAL EQUIPMENT(B)	438	.787	-	-	-
TURNKEY(C)	1814	.703	-	-	-
SMALL	-	-	1.617	3.951	2.443
LARGE	-	-	1.639	3.847	2.348
VENTURI-PACKED SCRUBBER HIGH EFFICIENCY					
COLLECTOR ONLY(A)	304	.784	-	-	-
TOTAL EQUIPMENT(B)	511	.775	-	-	-
TURNKEY(C)	1943	.701	-	-	-
SMALL	-	-	1.588	3.813	2.401
LARGE	-	-	1.581	3.667	2.319

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 235

DERIVED COST INDICES FOR DAP PROCESS PLANTS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
VENTURI SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	316	.716	-	-	-
TOTAL EQUIPMENT(B)	179	.870	-	-	-
TURNKEY(C)	2964	.579	-	-	-
SMALL	-	-	1.542	3.848	2.495
MEDIUM	-	-	1.641	3.641	2.218
VENTURI SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	377	.690	-	-	-
TOTAL EQUIPMENT(B)	269	.811	-	-	-
TURNKEY(C)	655	.798	-	-	-
MEDIUM	-	-	1.641	3.641	2.218
LARGE	-	-	1.737	3.829	2.204

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 236

DERIVED COST INDICES FOR DAP PROCESS PLANTS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
2-STAGE CYC. SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	360	.755	-	-	-
TOTAL EQUIPMENT(B)	282	.832	-	-	-
TURNKEY(C)	5399	.516	-	-	-
SMALL	-	-	1.287	3.163	2.457
MEDIUM	-	-	1.328	2.870	2.162
2-STAGE CYC. SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	480	.714	-	-	-
TOTAL EQUIPMENT(B)	424	.773	-	-	-
TURNKEY(C)	995	.761	-	-	-
MEDIUM	-	-	1.328	2.870	2.162
LARGE	-	-	1.365	2.934	2.150

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 237

DERIVED COST INDICES FOR DAP PROCESS PLANTS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
PACKED C-F SCRUBBERS MED. EFFICIENCY					
COLLECTOR ONLY(A)	295	.656	-	-	-
TOTAL EQUIPMENT(B)	256	.735	-	-	-
TURNKEY(C)	1509	.574	-	-	-
SMALL	-	-	1.452	3.000	2.066
MEDIUM	-	-	1.499	2.901	1.936
PACKED C-F SCRUBBER HIGH EFFICIENCY					
COLLECTOR ONLY(A)	140	.789	-	-	-
TOTAL EQUIPMENT(B)	158	.821	-	-	-
TURNKEY(C)	1151	.623	-	-	-
SMALL	-	-	1.404	2.808	2.000
MEDIUM	-	-	1.423	2.626	1.846

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 238

DERIVED COST INDICES FOR DAP PROCESS PLANTS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
PACKED C-F SCRUBBER MED. EFFICIENCY					
COLLECTOR ONLY(A)	35	.967	-	-	-
TOTAL EQUIPMENT(B)	135	.828	-	-	-
TURNKEY(C)	595	.709	-	-	-
MEDIUM	-	-	1.499	2.901	1.936
LARGE	-	-	1.404	2.570	1.830
PACKED C-F SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	133	.795	-	-	-
TOTAL EQUIPMENT(B)	325	.717	-	-	-
TURNKEY(C)	957	.650	-	-	-
MEDIUM	-	-	1.423	2.626	1.846
LARGE	-	-	1.371	2.452	1.788

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 239

DERIVED COST INDICES FOR FLOUR MILLING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
FABRIC FILTERS					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	15	.626	-	-	-
TOTAL EQUIPMENT(B)	21	.628	-	-	-
TURNKEY(C)	30	.622	-	-	-
SMALL	-	-	1.415	1.903	1.345
LARGE	-	-	1.420	1.896	1.335

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 240

DERIVED COST INDICES FOR FEED GRINDING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
FABRIC FILTERS					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	1395	.734	-	-	-
TOTAL EQUIPMENT(B)	1761	.713	-	-	-
TURNKEY(C)	3553	.692	-	-	-
SMALL	-	-	1.215	2.361	1.943
LARGE	-	-	1.180	2.227	1.887

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 241

DERIVED COST INDICES FOR FEED FLASH DRYERS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
THERMAL INCINERATOR					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	187	.648	-	-	-
TOTAL EQUIPMENT(B)	176	.662	-	-	-
TURNKEY(C)	279	.635	-	-	-
SMALL	-	-	1.065	1.336	1.255
LARGE	-	-	1.094	1.304	1.192

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 242

DERIVED COST INDICES FOR FEED FLASH DRYERS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
WET SCRUBBER MED. EFFICIENCY					
COLLECTOR ONLY(A)	44	.611	-	-	-
TOTAL EQUIPMENT(B)	77	.577	-	-	-
TURNKEY(C)	570	.443	-	-	-
SMALL	-	-	1.293	2.915	2.255
LARGE	-	-	1.210	2.105	1.740
WET SCRUBBER HIGH EFFICIENCY					
COLLECTOR ONLY(A)	41	.622	-	-	-
TOTAL EQUIPMENT(B)	72	.587	-	-	-
TURNKEY(C)	538	.451	-	-	-
SMALL	-	-	1.282	2.845	2.220
LARGE	-	-	1.198	2.042	1.705

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 243

DERIVED COST INDICES FOR OPEN KETTLES

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
INCINERATORS W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	6473	.118	-	-	-
TOTAL EQUIPMENT(B)	6481	.142	-	-	-
TURNKEY(C)	12443	.120	-	-	-
SMALL	-	-	1.136	1.941	1.709
LARGE	-	-	1.184	1.948	1.645

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 244

DERIVED COST INDICES FOR OPEN KETTLES

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
CAT. INCRS. W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	1619	.378	-	-	-
TOTAL EQUIPMENT(B)	1978	.372	-	-	-
TURNKEY(C)	4805	.297	-	-	-
SMALL	-	-	1.185	1.937	1.634
LARGE	-	-	1.174	1.686	1.436

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 245

DERIVED COST INDICES FOR RESIN REACTOR

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
INCINERATORS W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	336	.457	-	-	-
TOTAL EQUIPMENT(B)	322	.482	-	-	-
TURNKEY(C)	972	.408	-	-	-
SMALL	-	-	1.184	1.932	1.632
LARGE	-	-	1.224	1.811	1.480
INCRS. WITH H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	1682	.395	-	-	-
TOTAL EQUIPMENT(B)	1352	.367	-	-	-
TURNKEY(C)	2432	.351	-	-	-
SMALL	-	-	1.100	1.508	1.371
LARGE	-	-	1.132	1.520	1.342

FOR USE IN EQUATION COST = K(SIZE)**EXP(X)

TABLE 246

DERIVED COST INDICES FOR RESIN REACTORS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
CAT. INCRS. W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	825	.478	-	-	-
TOTAL EQUIPMENT(B)	451	.483	-	-	-
TURNKEY(C)	889	.454	-	-	-
SMALL	-	-	1.170	1.669	1.422
LARGE	-	-	1.197	1.616	1.350
CAT. INCRS. WITH H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	1130	.402	-	-	-
TOTAL EQUIPMENT(B)	1023	.430	-	-	-
TURNKEY(C)	1725	.401	-	-	-
SMALL	-	-	1.145	1.510	1.318
LARGE	-	-	1.190	1.508	1.268

FOR USE IN EQUATION COST = K(SIZE)**EXP(X)

TABLE 247

DERIVED COST INDICES FOR WEB-OFFSET PRINTING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
INCINERATORS W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	3153	.214	-	-	-
TOTAL EQUIPMENT(B)	2751	.247	-	-	-
TURNKEY(C)	6347	.205	-	-	-
SMALL	-	-	1.116	1.871	1.676
LARGE	-	-	1.163	1.848	1.590
CAT. INCRS. W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	354	.516	-	-	-
TOTAL EQUIPMENT(B)	373	.522	-	-	-
TURNKEY(C)	1137	.427	-	-	-
SMALL	-	-	1.100	1.628	1.480
LARGE	-	-	1.108	1.456	1.314

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 248

DERIVED COST INDICES FOR METAL DECORATING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
INCINERATOR W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	173	.551	-	-	-
TOTAL EQUIPMENT(B)	191	.555	-	-	-
TURNKEY(C)	1071	.426	-	-	-
SMALL	-	-	1.139	2.186	1.918
LARGE	-	-	1.144	1.949	1.704
INCRS. WITH H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	5465	.215	-	-	-
TOTAL EQUIPMENT(B)	3713	.271	-	-	-
TURNKEY(C)	6488	.258	-	-	-
SMALL	-	-	1.086	1.690	1.556
LARGE	-	-	1.144	1.757	1.537

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 249

DERIVED COST INDICES FOR METAL DECORATING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
CAT. INCRS. W/O H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	302	.537	-	-	-
TOTAL EQUIPMENT(B)	259	.566	-	-	-
TURNKEY(C)	542	.515	-	-	-
SMALL	-	-	1.087	1.493	1.374
LARGE	-	-	1.116	1.463	1.311
CAT. INCRS. WITH H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	695	.476	-	-	-
TOTAL EQUIPMENT(B)	570	.510	-	-	-
TURNKEY(C)	1012	.473	-	-	-
SMALL	-	-	1.084	1.414	1.304
LARGE	-	-	1.119	1.409	1.260

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 250

DERIVED COST INDICES FOR GRAVURE PRINTING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
INCRS. WITH H.E.					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	164	.604	-	-	-
TOTAL EQUIPMENT(B)	114	.653	-	-	-
TURNKEY(C)	324	.579	-	-	-
SMALL	-	-	1.108	1.560	1.408
LARGE	-	-	1.146	1.533	1.337

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 251

DERIVED COST INDICES FOR GRAVURE PRINTING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
CARBON ADSORPTION					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	73398	.776	-	-	-
TOTAL EQUIPMENT(B)	77154	.772	-	-	-
TURNKEY(C)	159821	.701	-	-	-
SMALL	-	-	1.046	1.964	1.878
LARGE	-	-	1.042	1.810	1.738

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 252

DERIVED COST INDICES FOR GRAY IRON CUPOLAS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
FABRIC FILTERS					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	4860	.738	-	-	-
TOTAL EQUIPMENT(B)	9865	.762	-	-	-
TURNKEY(C)	25195	.706	-	-	-
SMALL	-	-	2.142	4.808	2.245
LARGE	-	-	2.233	4.535	2.031

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 253

DERIVED COST INDICES FOR GRAY IRON CUPOLAS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
WET SCRUBBERS LA PROCESS WT.					
COLLECTOR ONLY(A)	4793	.600	-	-	-
TOTAL EQUIPMENT(B)	27663	.438	-	-	-
TURNKEY(C)	83726	.363	-	-	-
SMALL	-	-	3.974	10.119	2.546
LARGE	-	-	2.972	6.617	2.226
WET SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	5367	.624	-	-	-
TOTAL EQUIPMENT(B)	32093	.469	-	-	-
TURNKEY(C)	90839	.395	-	-	-
SMALL	-	-	4.183	10.003	2.391
LARGE	-	-	3.168	6.643	2.097

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 254

DERIVED COST INDICES FOR ROTARY LIME SLUDGE KILNS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
ELCSTC. PRCPTRS. LA PROCESS WT.					
COLLECTOR ONLY(A)	146	.593	-	-	-
TOTAL EQUIPMENT(B)	278	.562	-	-	-
TURNKEY(C)	332	.593	-	-	-
SMALL	-	-	1.386	2.283	1.647
LARGE	-	-	1.336	2.284	1.709
ELCSTC. PRCPTRS. HIGH EFFICIENCY					
COLLECTOR ONLY(A)	225	.568	-	-	-
TOTAL EQUIPMENT(B)	484	.525	-	-	-
TURNKEY(C)	781	.529	-	-	-
SMALL	-	-	1.372	2.305	1.679
LARGE	-	-	1.303	2.198	1.687

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 255

DERIVED COST INDICES FOR ROTARY LIME SLUDGE KILNS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
WET SCRUBBERS LA PROCESS WT.					
COLLECTOR ONLY(A)	31	.601	-	-	-
TOTAL EQUIPMENT(B)	17	.715	-	-	-
TURNKEY(C)	326	.524	-	-	-
SMALL	-	-	1.780	4.695	2.637
LARGE	-	-	2.044	4.280	2.094
WET SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	51	.570	-	-	-
TOTAL EQUIPMENT(B)	26	.697	-	-	-
TURNKEY(C)	402	.522	-	-	-
SMALL	-	-	1.915	4.765	2.488
LARGE	-	-	2.231	4.494	2.014

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 256

DERIVED COST INDICES FOR VERTICAL LIME ROCK KILNS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
ELCSTC. PRCPTRS. LA PROCESS WT.					
COLLECTOR ONLY(A)	401	.450	-	-	-
TOTAL EQUIPMENT(B)	311	.508	-	-	-
TURNKEY(C)	339	.543	-	-	-
SMALL	-	-	1.450	2.306	1.591
LARGE	-	-	1.509	2.459	1.629
ELCSTC. PRCPTRS. HIGH EFFICIENCY					
COLLECTOR ONLY(A)	315	.496	-	-	-
TOTAL EQUIPMENT(B)	352	.515	-	-	-
TURNKEY(C)	742	.492	-	-	-
SMALL	-	-	1.381	2.268	1.642
LARGE	-	-	1.400	2.262	1.616

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 257

DERIVED COST INDICES FOR VERTICAL LIME ROCK KILNS

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
FABRIC FILTERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	19	.770	-	-	-
TOTAL EQUIPMENT(B)	68	.665	-	-	-
TURNKEY(C)	238	.592	-	-	-
SMALL	-	-	1.340	2.297	1.713
LARGE	-	-	1.179	1.846	1.566

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 258

DERIVED COST INDICES FOR VERTICAL LIME ROCK KILN

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
WET SCRUBBERS LA PROCESS WT.					
COLLECTOR ONLY(A)	34	.551	-	-	-
TOTAL EQUIPMENT(B)	42	.579	-	-	-
TURNKEY(C)	476	.443	-	-	-
SMALL	-	-	1.658	5.056	3.050
LARGE	-	-	1.717	4.435	2.583
WET SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	34	.550	-	-	-
TOTAL EQUIPMENT(B)	101	.508	-	-	-
TURNKEY(C)	800	.403	-	-	-
SMALL	-	-	1.963	5.647	2.877
LARGE	-	-	1.864	4.716	2.529

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

TABLE 239

DERIVED COST INDICES FOR DETERGENT SPRAY DRYING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
WET SCRUBBERS MED. EFFICIENCY					
COLLECTOR ONLY(A)	30	.641	-	-	-
TOTAL EQUIPMENT(B)	73	.696	-	-	-
TURNKEY(C)	485	.552	-	-	-
SMALL	-	-	1.482	3.338	2.238
LARGE	-	-	1.492	2.814	1.953
WET SCRUBBERS HIGH EFFICIENCY					
COLLECTOR ONLY(A)	83	.583	-	-	-
TOTAL EQUIPMENT(B)	126	.603	-	-	-
TURNKEY(C)	631	.535	-	-	-
SMALL	-	-	1.706	4.230	2.338
LARGE	-	-	1.856	3.843	2.070

FOR USE IN EQUATION COST = K(SIZE)**EXP(X)

TABLE 260

DERIVED COST INDICES FOR DETERGENT SPRAY DRYING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
FABRIC FILTERS					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	2	.385	-	-	-
TOTAL EQUIPMENT(B)	5	.428	-	-	-
TURNKEY(C)	10	.515	-	-	-
SMALL	-	-	1.320	2.460	1.860
LARGE	-	-	1.226	2.242	1.828

FOR USE IN EQUATION COST = K(SIZE)**EXP(X*)

TABLE 261

DERIVED COST INDICES FOR SOAP & DET. PROD. HANDLING

COLLECTOR TYPE	K*	X*	B/A	C/A	C/B
FABRIC FILTERS					
HIGH EFFICIENCY					
COLLECTOR ONLY(A)	1	.999	-	-	-
TOTAL EQUIPMENT(B)	3	.942	-	-	-
TURNKEY(C)	6	.926	-	-	-
SMALL	-	-	1.350	2.516	1.864
LARGE	-	-	1.246	2.272	1.824

FOR USE IN EQUATION COST = K(SIZE)*EXP(X)

Equipment Type	Maximum Value	Minimum Value	Arithmetic Average
Fabric Filters			
Collector Only	.999	.626	.816
Turnkey System	.915	.622	.772
Carbon Adsorption			
Collector Only	–	–	.776
Turnkey System	–	–	.701
Wet Scrubbers			
Collector Only	.795	.248	.626
Turnkey System	.798	.200	.514
Electrostatic Precipitators			
Collector Only	.593	.450	.527
Turnkey System	.598	.492	.539
Incinerators			
Collector Only	.648	.118	.402
Turnkey System	.635	.120	.396

Of the five types of equipment involved, fabric filters had the highest exponent. This was to be expected since fabric filtering equipment tends to be additive. That is, to increase the capacity of a fabric filter system one can many times add another unit. This becomes more practicable with larger air flows. Many of the fabric filter systems were under 25,000 ACFM. Had more been in the 50,000-and-up range, the cost-to-size relationship would have been even closer to linear.

The exponent for incinerators had values from .120 in the 500 ACFM range to .635 in the 100,000 ACFM range. The average was 0.4 – the lowest for all the types of control systems considered in this report. This was below the 0.6 exponent usually assumed, because a number of the incinerators were of relatively small size. Every incinerator, regardless of size, requires an extensive system of safety control devices to prevent explosions. This system is a significant part of the cost of incinerators. So in the small size range the cost of increasing size is relatively small.

The average exponent derived for carbon adsorption collectors was .776. The limited amount of data on these systems precludes drawing many conclusions, but the cost data presented in this report could be safely extrapolatable down to one gravure press, or 12,450 ACFM.

Wet scrubbers had an average exponent of .626 – somewhat higher than the 0.6 usually assumed for equipment. A number of the scrubbers had an inlet gas flow rate above 100,000 ACFM. In this range the basic design of scrubbers must be changed to handle the high volumes of gas, and the cost starts increasing more rapidly with size.

Exponents for electrostatic precipitators varied from .450 to .593, with an average value of .527. In only one application was the gas flow above 100,000 ACFM. Precipitator flow rates from 300,000 to 600,000 ACFM are not unusual. A major capital expense of precipitators is the power supply that is required. This costs nearly as much for small precipitators as for large ones. Therefore, the cost of small precipitators does not increase as rapidly with size as would be expected in larger designs.

The use of the derived capital cost equations outside the range of the data from which they were calculated is valid within certain limitations. Very small equipment installations tend to have relatively high capital costs which do not correlate well with size. Small systems cost roughly the same regardless of the treated gas throughput. Very large systems are frequently based on different designs than their smaller counterparts, or are composed of several smaller units. Consequently, the derived cost indices will be inaccurate for these larger sizes. Numerical values for these large and small limitations depend upon both the nature of the abatement equipment and the nature of the process to which it is applied. Generalizations of these numerical values can be made, however, and they are presented below as guidelines.

	Small Limit, ACFM	Large Limit, ACFM
Scrubbers	2,000	100,000
Fabric Filters	2,000	very large
Precipitators	50,000	very large
Incinerators	2,000	50,000

The basic capital cost data collected were also used to calculate the cost per SCFM of inlet gas for each application. Results of these calculations are presented in the following tables:

Process Area	Table Numbers
Phosphates	
– WPPA	262
– SPA	263
– GTSP	264
– DAP	265, 266, 267, 268
Feed and Grain	269, 270, 271, 272
Paint and Varnish	273, 274, 275, 276
Graphic Arts	277, 278, 279, 280, 281
Gray Iron Foundries	282, 283
Lime Kilns	284, 285, 286, 287, 288
Soap and Detergent	289, 290, 291

TABLE 262

DERIVED COST PER SCFM* FOR WPPA PROCESS PLANTS

COLLECTOR TYPE	SMALL	LARGE
PACKED C-F SCRUBBERS		
ONE-THIRD GAS FLOW		
GAS FLOW, SCFM	8332	11954
COLLECTOR ONLY	1.00	.81
TOTAL EQUIPMENT	1.53	1.25
TURNKEY SYSTEM	4.63	3.63
WET SCRUBBERS		
NORMAL GAS FLOW		
GAS FLOW, SCFM	8332	11954
COLLECTOR ONLY	2.12	1.80
TOTAL EQUIPMENT	3.15	2.60
TURNKEY SYSTEM	7.50	6.02

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 263

DERIVED COST PER SCFM* FOR SPA PROCESS PLANTS

COLLECTOR TYPE	SMALL	LARGE
VENTURI SCRUBBERS		
NOMINAL EFFICIENCY		
GAS FLOW,SCFM	1252	2506
COLLECTOR ONLY	5.10	3.43
TOTAL EQUIPMENT	5.10	3.43
TURNKEY SYSTEM	7.33	4.73
VENTURI-PACKED SCRUBBER		
NOMINAL EFFICIENCY		
GAS FLOW,SCFM	1252	2506
COLLECTOR ONLY	5.02	3.09
TOTAL EQUIPMENT	8.35	4.96
TURNKEY SYSTEM	19.21	11.41

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 264

DERIVED COST PER SCFM* FOR GTSP PROCESS PLANTS

COLLECTOR TYPE	SMALL	LARGE
VENTURI-PACKED SCRUBBER		
MOD. EFFICIENCY		
GAS FLOW, SCFM	33870	54077
COLLECTOR ONLY	1.09	.98
TOTAL EQUIPMENT	1.76	1.60
TURNKEY SYSTEM	4.31	3.76
VENTURI-PACKED SCRUBBER		
HIGH EFFICIENCY		
GAS FLOW, SCFM	33870	54077
COLLECTOR ONLY	1.20	1.09
TOTAL EQUIPMENT	1.91	1.72
TURNKEY SYSTEM	4.58	3.99

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 265

DERIVED COST PER SCFM FOR DAP PROCESS PLANTS

COLLECTOR TYPE		
VENTURI SCRUBBERS HIGH EFFICIENCY	(SMALL)	(MEDIUM)
GAS FLOW,SCFM	30641	46127
COLLECTOR ONLY	1.09	.96
TOTAL EQUIPMENT	1.67	1.58
TURNKEY SYSTEM	4.18	3.51
VENTURI SCRUBBERS HIGH EFFICIENCY	(MEDIUM)	(LARGE)
GAS FLOW,SCFM	46127	73703
COLLECTOR ONLY	.96	.83
TOTAL EQUIPMENT	1.58	1.45
TURNKEY SYSTEM	3.51	3.19

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 266

DERIVED COST PER SCFM FOR DAP PROCESS PLANTS

COLLECTOR TYPE		
CYCLONIC SCRUBBERS		
LA PROCESS WT.	(SMALL)	(MEDIUM)
GAS FLOW,SCFM	30641	46127
COLLECTOR ONLY	1.60	1.44
TOTAL EQUIPMENT	2.05	1.91
TURNKEY SYSTEM	5.05	4.13
CYCLONIC SCRUBBERS		
LA PROCESS WT.	(MEDIUM)	(LARGE)
GAS FLOW,SCFM	46127	73703
COLLECTOR ONLY	1.44	1.26
TOTAL EQUIPMENT	1.91	1.72
TURNKEY SYSTEM	4.13	3.70

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 267

DERIVED COST PER SCFM FOR DAP PROCESS PLANTS

COLLECTOR TYPE	SMALL	MEDIUM
PACKED C-F SCRUBBERS MED. EFFICIENCY		
GAS FLOW,SCFM	30227	44305
COLLECTOR ONLY	.70	.62
TOTAL EQUIPMENT	1.01	.93
TURNKEY SYSTEM	2.09	1.80
PACKED C-F SCRUBBER HIGH EFFICIENCY		
GAS FLOW,SCFM	30227	44305
COLLECTOR ONLY	.78	.73
TOTAL EQUIPMENT	1.09	1.04
TURNKEY SYSTEM	2.19	1.92

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 268

DERIVED COST PER SCFM FOR DAP PROCESS PLANTS

COLLECTOR TYPE	MEDIUM	LARGE
PACKED C-F SCRUBBER		
MED. EFFICIENCY		
GAS FLOW,SCFM	44305	70805
COLLECTOR ONLY	.62	.61
TOTAL EQUIPMENT	.93	.86
TURNKEY SYSTEM	1.80	1.57
PACKED C-F SCRUBBERS		
HIGH EFFICIENCY		
GAS FLOW,SCFM	44305	70805
COLLECTOR ONLY	.73	.66
TOTAL EQUIPMENT	1.04	.91
TURNKEY SYSTEM	1.92	1.63

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 269

DERIVED COST PER SCFM FOR FLOUR MILLING

COLLECTOR TYPE	SMALL	LARGE
FABRIC FILTERS HIGH EFFICIENCY		
GAS FLOW,SCFM	6000	20000
COLLECTOR ONLY	.92	.58
TOTAL EQUIPMENT	1.11	.83
TURNKEY SYSTEM	1.76	1.17

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR
INLET INCLUDING WATER VAPOR

TABLE 270

DERIVED COST PER SCFM FOR FEED GRINDING

COLLECTOR TYPE	SMALL	LARGE
FABRIC FILTERS HIGH EFFICIENCY GAS FLOW, SCFM* COLLECTOR ONLY TOTAL EQUIPMENT TURNKEY SYSTEM	 8000 1.27 1.92 5.76	 27000 .90 .98 3.78

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR
INLET INCLUDING WATER VAPOR

TABLE 271

DERIVED COST PER SCFM FOR FEED FLASH DRYERS

COLLECTOR TYPE	SMALL	LARGE
THERMAL INCINERATOR HIGH EFFICIENCY GAS FLOW,SCFM COLLECTOR ONLY TOTAL EQUIPMENT TURNKEY SYSTEM	 8013 7.52 7.18 9.53	 18000 5.18 5.85 6.67

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR
INLET INCLUDING WATER VAPOR

TABLE 272

DERIVED COST PER SCFM FOR FEED FLASH DRYERS

COLLECTOR TYPE	SMALL	LARGE
WET SCRUBBER		
MED. EFFICIENCY		
GAS FLOW,SCFM	8413	58889
COLLECTOR ONLY	1.20	.56
TOTAL EQUIPMENT	1.55	.68
TURNKEY SYSTEM	3.50	1.18
WET SCRUBBER		
HIGH EFFICIENCY		
GAS FLOW,SCFM	8413	58889
COLLECTOR ONLY	1.25	.60
TOTAL EQUIPMENT	1.60	.71
TURNKEY SYSTEM	3.54	1.22

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 273

DERIVED COST PER SCFM FOR OPEN KETTLES

COLLECTOR TYPE	SMALL	LARGE
INCINERATORS W/O H.E. HIGH EFFICIENCY		
GAS FLOW,SCFM	500	2999
COLLECTOR ONLY	24.18	4.94
TOTAL EQUIPMENT	27.47	5.84
TURNKEY SYSTEM	46.95	9.61

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 274

DERIVED COST PER SCFM FOR OPEN KETTLES

COLLECTOR TYPE	SMALL	LARGE
CAT. INCRS. W/O H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	500	2999
COLLECTOR ONLY	23.97	7.66
TOTAL EQUIPMENT	28.41	8.99
TURNKEY SYSTEM	46.43	12.92

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 275

DERIVED COST PER SCFM FOR RESIN REACTORS

COLLECTOR TYPE	SMALL	LARGE
INCINERATORS W/O H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	2999	9996
COLLECTOR ONLY	4.94	2.71
TOTAL EQUIPMENT	5.84	3.31
TURNKEY SYSTEM	9.54	4.90
INCRS. WITH H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	2999	9996
COLLECTOR ONLY	9.90	4.69
TOTAL EQUIPMENT	10.88	5.31
TURNKEY SYSTEM	14.92	7.12

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 276

DERIVED COST PER SCFM FOR RESIN REACTORS

COLLECTOR TYPE	SMALL	LARGE
CAT. INCRS. W/O H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	2999	9996
COLLECTOR ONLY	7.67	4.33
TOTAL EQUIPMENT	9.00	5.18
TURNKEY SYSTEM	12.79	6.99
CAT. INCRS. WITH H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	2999	9996
COLLECTOR ONLY	10.57	5.39
TOTAL EQUIPMENT	12.11	6.41
TURNKEY SYSTEM	15.96	8.13

BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 277

DERIVED COST PER SCFM FOR WEB-OFFSET PRINTING

COLLECTOR TYPE	SMALL	LARGE
INCINERATORS W/O H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	2002	7001
COLLECTOR ONLY	8.03	3.00
TOTAL EQUIPMENT	8.96	3.49
TURNKEY SYSTEM	15.02	5.55
CAT. INCRS. W/O H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	2002	7001
COLLECTOR ONLY	8.97	4.90
TOTAL EQUIPMENT	9.86	5.43
TURNKEY SYSTEM	14.60	7.13

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 278

DERIVED COST PER SCFM FOR METAL DECORATING

COLLECTOR TYPE	SMALL	LARGE
INCINERATOR W/O H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	3991	9978
COLLECTOR ONLY	4.20	2.79
TOTAL EQUIPMENT	4.79	3.19
TURNKEY SYSTEM	9.19	5.43
INCRS. WITH H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	3991	9978
COLLECTOR ONLY	8.14	3.97
TOTAL EQUIPMENT	8.84	4.54
TURNKEY SYSTEM	13.76	6.97

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 279

DERIVED COST PER SCFM FOR METAL DECORATING

COLLECTOR TYPE	SMALL	LARGE
CAT. INCRS. W/O H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	3991	9978
COLLECTOR ONLY	6.52	4.27
TOTAL EQUIPMENT	7.09	4.76
TURNKEY SYSTEM	9.73	6.24
CAT. INCRS. WITH H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	3991	9978
COLLECTOR ONLY	9.06	5.61
TOTAL EQUIPMENT	9.82	6.27
TURNKEY SYSTEM	12.81	7.90

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 280

DERIVED COST PER SCFM FOR GRAVURE PRINTING

COLLECTOR TYPE	SMALL	LARGE
INCRS. WITH H.E.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	12045	24091
COLLECTOR ONLY	3.95	3.00
TOTAL EQUIPMENT	4.38	3.44
TURNKEY SYSTEM	6.16	4.60

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 281

DERIVED COST PER SCFM FOR GRAVURE PRINTING

COLLECTOR TYPE	SMALL	LARGE
CARBON ADSORPTION HIGH EFFICIENCY GAS FLOW,SCFM COLLECTOR ONLY TOTAL EQUIPMENT TURNKEY SYSTEM	 47989 4.48 4.69 8.81	 143967 3.50 3.65 6.94

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 282

DERIVED COST PER SCFM FOR GRAY IRON CUPOLAS

COLLECTOR TYPE	SMALL	LARGE
FABRIC FILTERS HIGH EFFICIENCY		
GAS FLOW,SCFM	8098	47515
COLLECTOR ONLY	2.96	7.30
TOTAL EQUIPMENT	6.33	4.77
TURNKEY SYSTEM	14.22	9.53

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE [illegible]

DERIVED COST PER SCFM FOR GRAY IRON CUPOLAS

COLLECTOR TYPE	SMALL	LARGE
WET SCRUBBERS LA PROCESS WT.		
GAS FLOW, SCFM	[illegible]	[illegible]
COLLECTOR ONLY	[illegible]	[illegible]
TOTAL EQUIPMENT	[illegible]	[illegible]
TURNKEY SYSTEM	[illegible]	[illegible]
WET SCRUBBERS HIGH EFFICIENCY		
GAS FLOW, SCFM	[illegible]	[illegible]
COLLECTOR ONLY	[illegible]	[illegible]
TOTAL EQUIPMENT	[illegible]	[illegible]
TURNKEY SYSTEM	[illegible]	[illegible]

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR
INLET INCLUDING WATER VAPOR

TABLE [illegible]

DERIVED COST PER SCFM FOR VERTICAL LIME KILN KILNS

COLLECTOR TYPE	SMALL	LARGE
WET SCRUBBERS LO PRESSURE DP.		
GAS FLOW, SCFM	[illegible]	[illegible]
COLLECTOR ONLY	[illegible]	[illegible]
TOTAL EQUIPMENT	[illegible]	[illegible]
TURNKEY SYSTEM	[illegible]	[illegible]
WET SCRUBBERS HIGH EFFICIENCY		
GAS FLOW, SCFM	[illegible]	[illegible]
COLLECTOR ONLY	[illegible]	[illegible]
TOTAL EQUIPMENT	[illegible]	[illegible]
TURNKEY SYSTEM	[illegible]	[illegible]

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR
INLET INCLUDING WATER VAPOR

TABLE 285

DERIVED COST PER SCFM FOR VERTICAL LIME ROCK KILNS

COLLECTOR TYPE	SMALL	LARGE
FABRIC FILTERS		
HIGH EFFICIENCY		
GAS FLOW,SCFM	5204	18300
COLLECTOR ONLY	5.81	4.27
TOTAL EQUIPMENT	7.78	5.83
TURNKEY SYSTEM	13.34	7.88

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR
INLET INCLUDING WATER VAPOR

TABLE 286

DERIVED COST PER SCFM FOR VERTICAL LIME ROCK KILNS

COLLECTOR TYPE	SMALL	LARGE
ELCSTC. PRCPTRS.		
LA PROCESS WT.		
GAS FLOW,SCFM	18388	36703
COLLECTOR ONLY	2.87	1.96
TOTAL EQUIPMENT	4.16	2.96
TURNKEY SYSTEM	6.61	4.82
ELCSTC. PRCPTRS.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	18388	36703
COLLECTOR ONLY	3.68	2.60
TOTAL EQUIPMENT	5.09	3.64
TURNKEY SYSTEM	8.36	5.89

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 287

DERIVED COST PER SCFM FOR ROTARY LIME SLUDGE KILNS

COLLECTOR TYPE	SMALL	LARGE
WET SCRUBBERS		
LA PROCESS WT.		
GAS FLOW,SCFM	22247	78518
COLLECTOR ONLY	.70	.41
TOTAL EQUIPMENT	1.25	.84
TURNKEY SYSTEM	3.30	1.76
WET SCRUBBERS		
HIGH EFFICIENCY		
GAS FLOW,SCFM	22247	78518
COLLECTOR ONLY	.84	.47
TOTAL EQUIPMENT	1.60	1.05
TURNKEY SYSTEM	3.99	2.12

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 288

DERIVED COST PER SCFM FOR ROTARY LIME SLUDGE KILNS

COLLECTOR TYPE	SMALL	LARGE
ELCSTC. PRCPTRS.		
LA PROCESS WT.		
GAS FLOW,SCFM	22247	78518
COLLECTOR ONLY	3.02	1.75
TOTAL EQUIPMENT	4.19	2.34
TURNKEY SYSTEM	6.90	4.00
ELCSTC. PRCPTRS.		
HIGH EFFICIENCY		
GAS FLOW,SCFM	22247	78518
COLLECTOR ONLY	3.60	2.02
TOTAL EQUIPMENT	4.95	2.64
TURNKEY SYSTEM	8.31	4.45

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 289

DERIVED COST PER SCFM FOR DETERGENT SPRAY DRYING

COLLECTOR TYPE	SMALL	LARGE
WET SCRUBBERS MED. EFFICIENCY		
GAS FLOW,SCFM	23961	95835
COLLECTOR ONLY	1.19	.72
TOTAL EQUIPMENT	1.76	1.08
TURNKEY SYSTEM	3.95	2.06
WET SCRUBBERS HIGH EFFICIENCY		
GAS FLOW,SCFM	23961	95835
COLLECTOR ONLY	1.17	.65
TOTAL EQUIPMENT	2.10	1.21
TURNKEY SYSTEM	4.90	2.51

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 290

DERIVED COST PER SCFM FOR DETERGENT SPRAY DRYING

COLLECTOR TYPE	SMALL	LARGE
FABRIC FILTERS HIGH EFFICIENCY		
GAS FLOW,SCFM	23961	95835
COLLECTOR ONLY	1.47	1.44
TOTAL EQUIPMENT	1.95	1.77
TURNKEY SYSTEM	3.64	3.23

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

TABLE 291

DERIVED COST PER SCFM FOR SOAP & DET. PROD. HANDLING

COLLECTOR TYPE	SMALL	LARGE
FABRIC FILTERS		
HIGH EFFICIENCY		
GAS FLOW, SCFM	18929	75714
COLLECTOR ONLY	1.31	1.31
TOTAL EQUIPMENT	1.77	1.63
TURNKEY SYSTEM	3.29	2.97

*BASED ON SCFM AT 70 DEG. F AT COLLECTOR INLET INCLUDING WATER VAPOR

4. GENERALIZED COST DATA

A series of correlations were made to investigate the general relationship of the cost of equipment to the gas flow rate for the types of control systems discussed in this report. These correlations were made using the data presented in the previous section of this report. The points plotted on each graphical correlation are coded so that the process application which they represent can be identified. The same point symbol code was used in all of the graphs and it is fully explained in Table 292.

Scrubbers

Correlations were made for both the scrubber cost and for the total installed cost of the scrubber system. Figure 180 shows the cost of the scrubber alone. Although there is a wide scatter in the data, the results correlate fairly well when the complexity of the scrubber is considered. Those scrubbers that fall along the "High Complexity" line are units employing a combination of scrubbing methods, such as venturi scrubbers with a packed section, to remove gaseous contaminants in a highly corrosive environment. Those falling along the "Low Complexity" line are mostly for removing particulates in a relatively non-corrosive environment. Many of the scrubbers in the "Medium" area were of the packed cross-flow type.

Figure 181 shows the cost for installed scrubbing systems. The wide variability in the applications of the scrubbing systems has caused a very wide scatter in the data. The only generalizations that can be made are that those scrubbers having a high complexity and using high energy would fall along the top edge of the plot area. Those of low complexity and using low energy would fall along the bottom. Scrubbers having a pressure drop of 30 or more inches of water have been defined here as being of high energy. Scrubbers defined here as being of low energy have a pressure drop of 8 inches of water or less. The intermediate area contains systems of high energy and low complexity, or high complexity but low energy.

In both Figure 180 and Figure 181 the bottom part of the curves are fairly straight, because the basis of the scrubber and the system remains essentially the same from very small scrubbers to moderately large ones; and therefore, a good cost-to-size relationship exists. But over 50,000 ACFM the basic design must be changed to handle the increasing gas flows; and the cost starts increasing more rapidly with size, causing the lines to curve upward.

TABLE 292

PLOTTING SYMBOL KEY

Process Area	Symbol
Phosphates	
– WPPA	□
– SPA	◇
– GTSP	✕
– DAP	○
Feed and Grain	△
Paint and Varnish	⬯
Graphic Arts	⬡
Soap and Detergent	◺
Lime Kilns	◊
Gray Iron Foundries	⬦

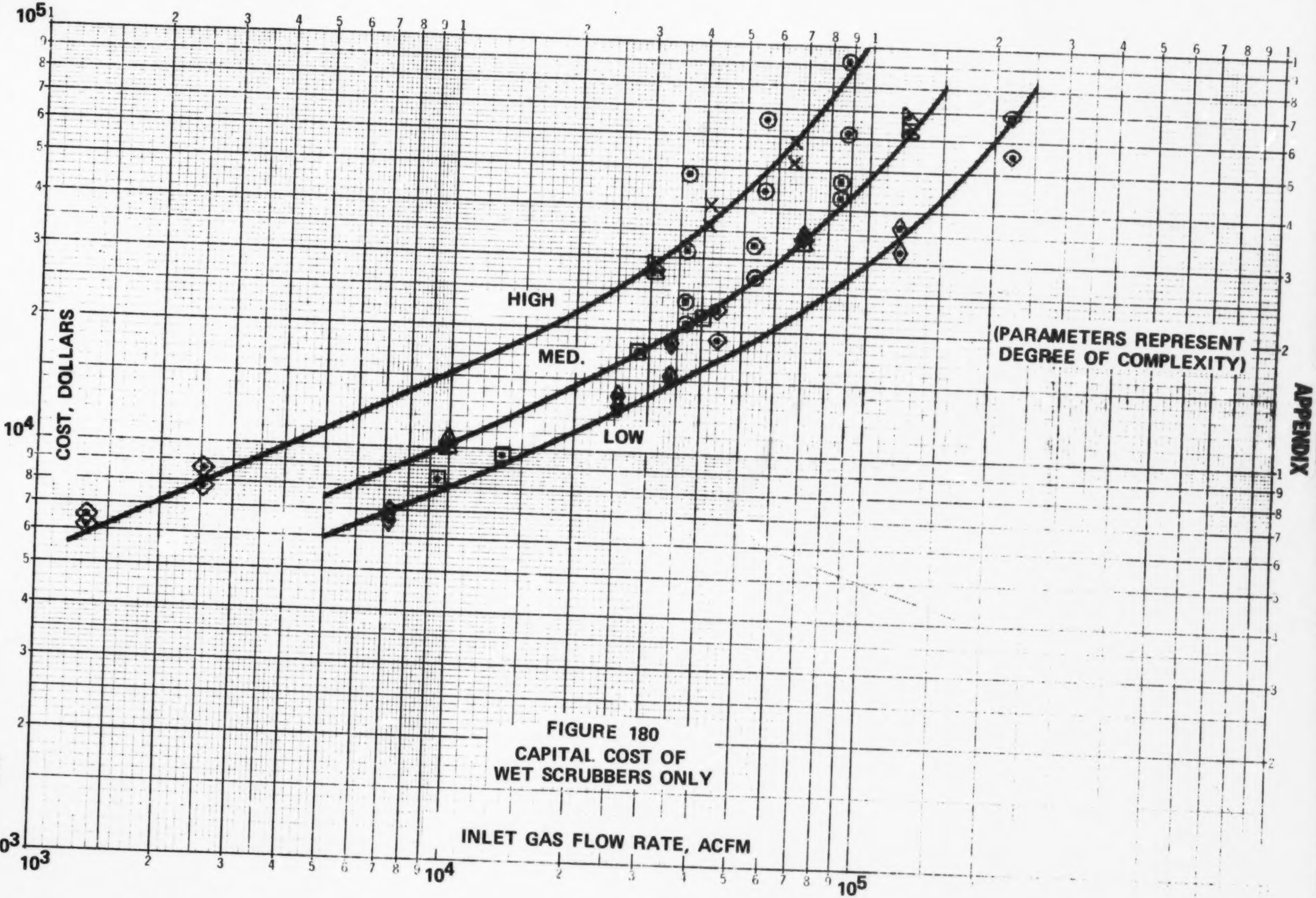


FIGURE 180
CAPITAL COST OF
WET SCRUBBERS ONLY

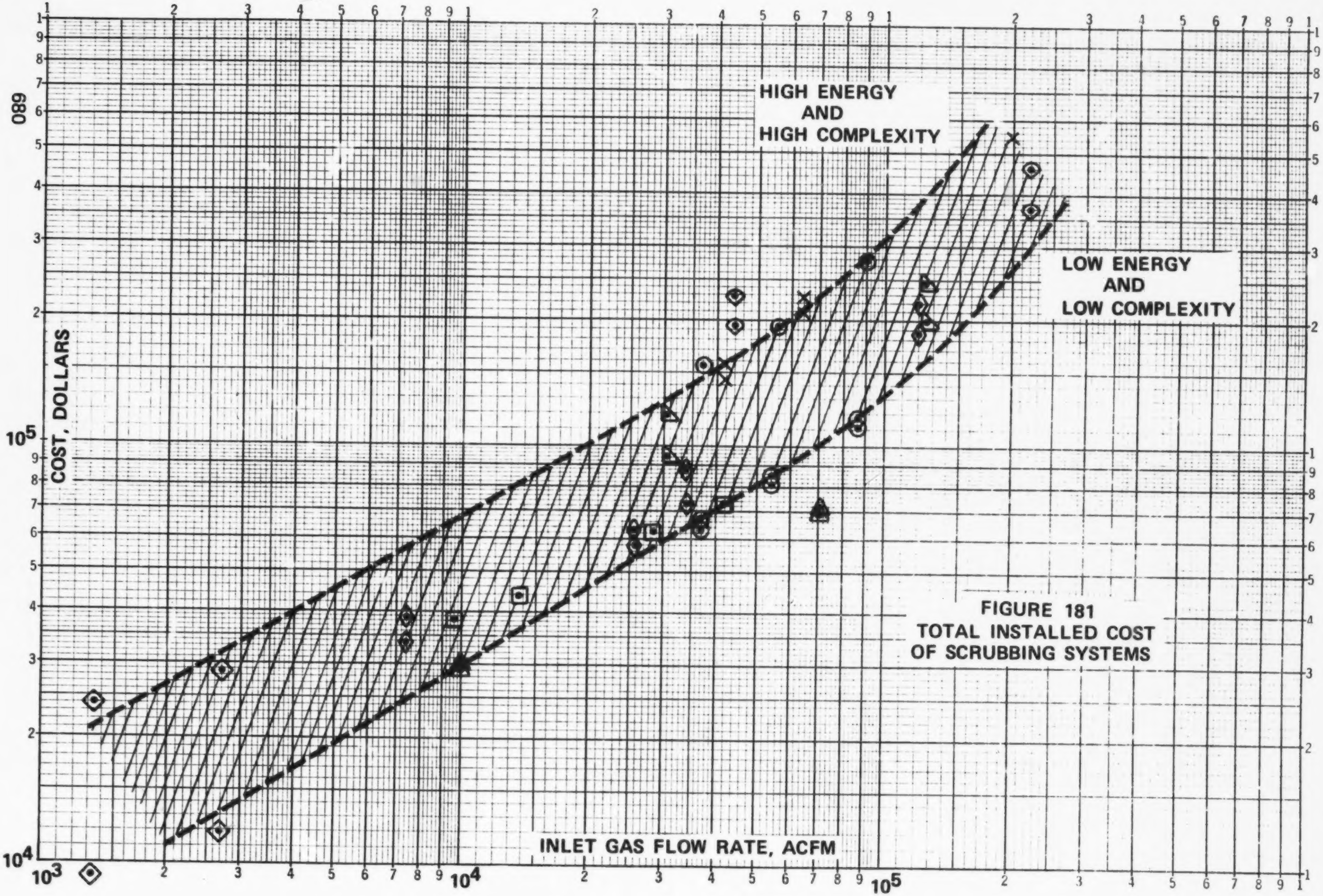


FIGURE 181
TOTAL INSTALLED COST
OF SCRUBBING SYSTEMS

The operating cost of scrubbers are shown in Figure 182. The results correlate well with the pressure drop across the system. The top line is for scrubbers with a pressure drop of 30 inches of water or greater, the middle for 10 to 16 inches, and the bottom for eight inches of water or less. At the top end of the top curve nearly all the operating costs are from the power for pushing the large volumes of gas through the scrubber, and the increase in operating cost is nearly linear with increasing gas flow rate. Each line starts curving up near its bottom end as the more fixed operating costs, such as operating and maintenance labor, become a more significant part of the total which, therefore, increases less rapidly with size.

Fabric Filters

Similar correlations were made for fabric filters. However, as Figure 183 illustrates, due to the wide variability in the applications involved, no true correlation exists between the different application on a gas flow rate basis. This is caused by the fact that the characteristics of the different particulates being collected require a different air-to-cloth ratio for each application. When the actual flow rate is divided by the air-to-cloth ratio, the result is the area, in square feet, of the fabric used. This area determines the physical size of the unit. As Figure 184 shows, when plotted against the fabric area, both the cost of the filter only and the installed system cost can be well represented by straight lines.

Figure 185 shows that when operating costs are plotted against the fabric area the results can be well represented by one line. As the size increases, the relationship between size and cost becomes nearly linear. This can be expected; for as the size increases, the cost of labor and material for replacing the bags become the major part of the operating cost. And these bag replacement costs are directly proportional to the size of the unit.

Incinerators

Basically, four types of incinerators were discussed in the previous section of this report. They are thermal incinerators with and without heat exchangers, and catalytic incinerators with and without heat exchangers.

Figure 186 shows the capital cost of incinerators without heat exchangers. In this, as well as in subsequent plots of capital costs, the curves are concave upward. Near the lower end of the curve, the cost does not increase with size

(Text continued on page 686)

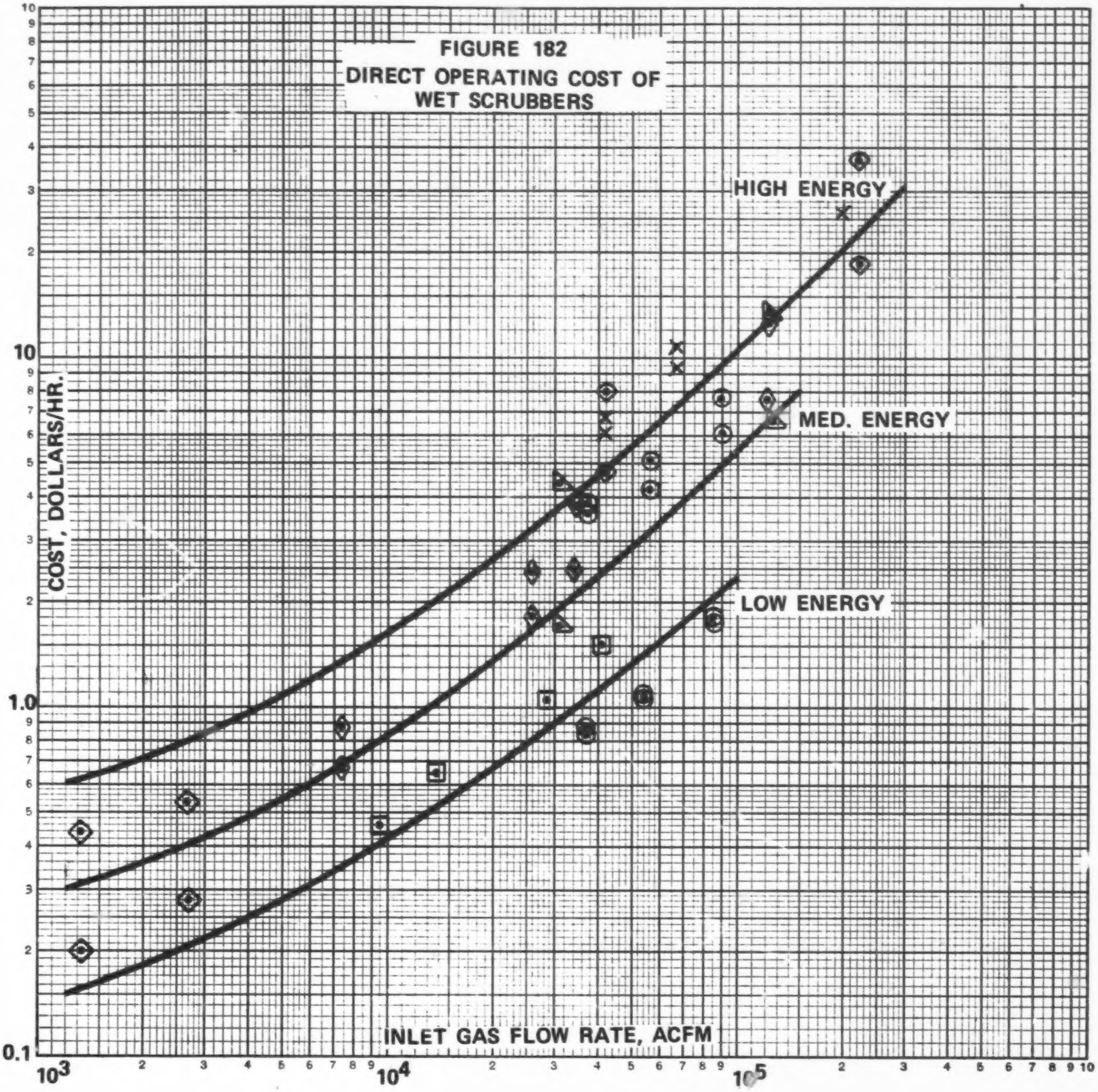


FIGURE 182
DIRECT OPERATING COST OF
WET SCRUBBERS
HIGH ENERGY
MED. ENERGY
LOW ENERGY
COST, DOLLARS/HR.
INLET GAS FLOW RATE, ACFM
0.1
1.0
10
10^3
10^4
10^5

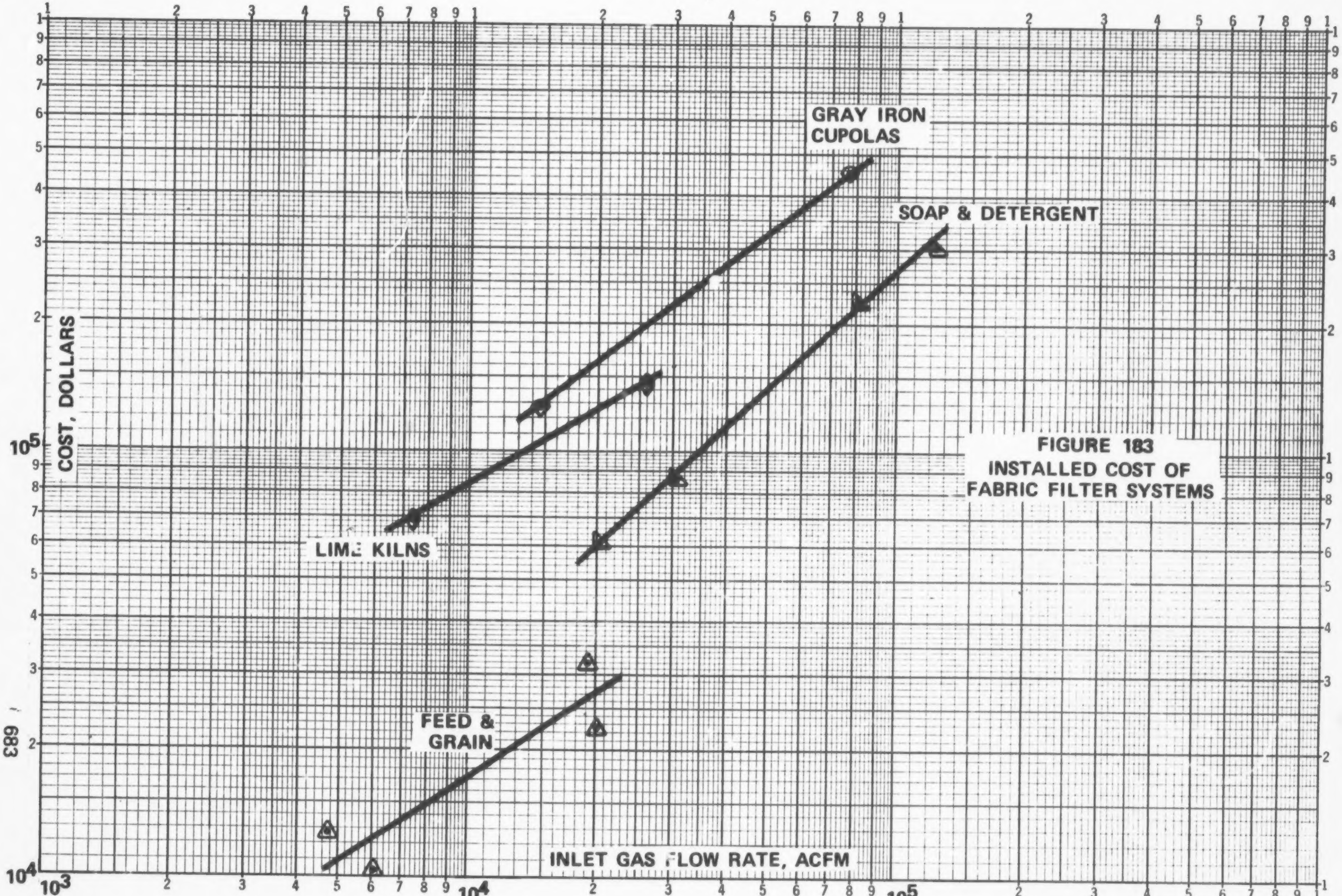


FIGURE 183
INSTALLED COST OF
FABRIC FILTER SYSTEMS

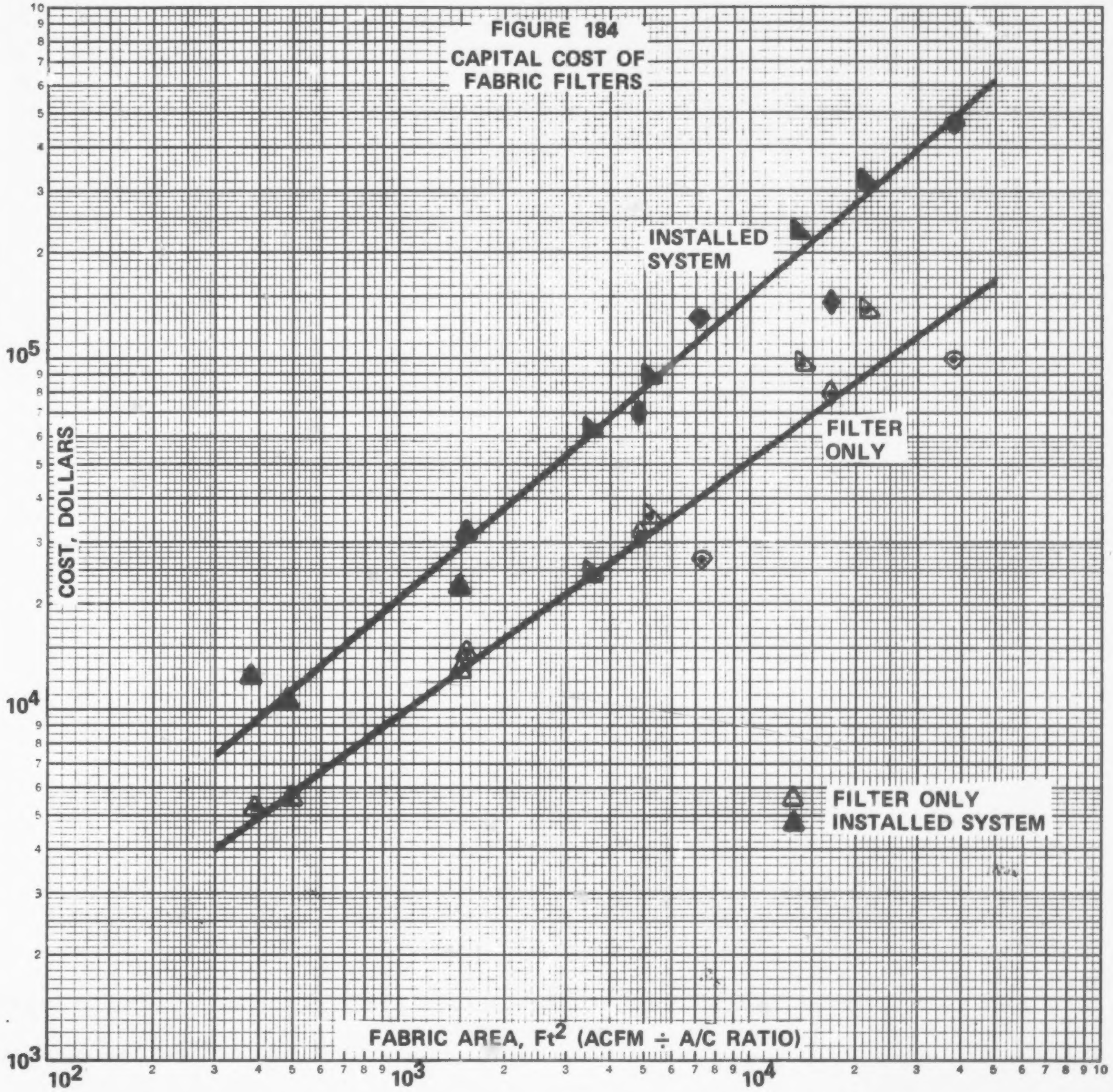


FIGURE 184
CAPITAL COST OF FABRIC FILTERS
INSTALLED SYSTEM
FILTER ONLY
△ FILTER ONLY
▲ INSTALLED SYSTEM
COST, DOLLARS
FABRIC AREA, Ft^2 (ACFM ÷ A/C RATIO)
10^3
10^4
10^5
10^2
2
3
4
5
6
7
8
9
10

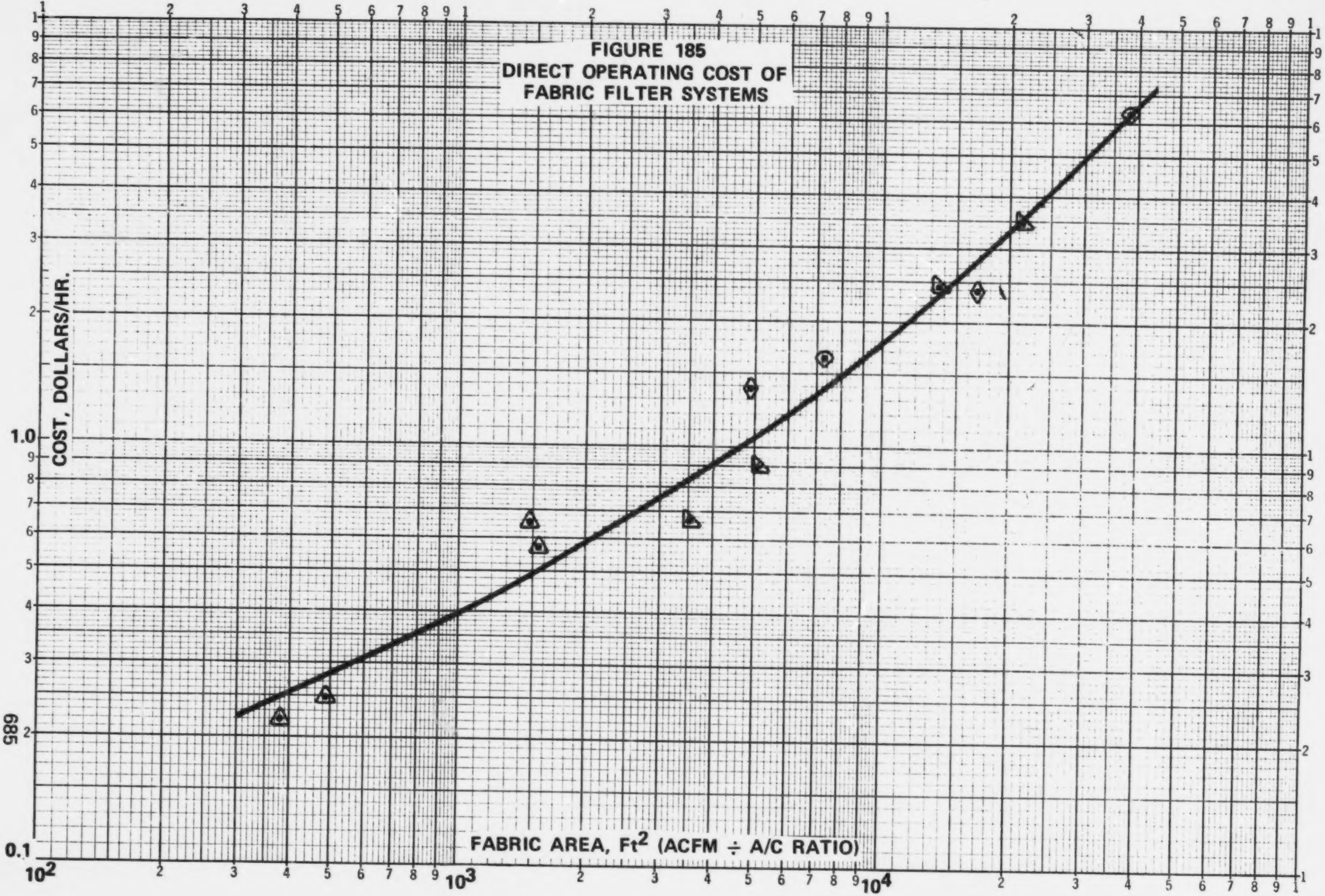


FIGURE 185
DIRECT OPERATING COST OF
FABRIC FILTER SYSTEMS

nearly as rapidly as it does toward the curve's upper end. This is due mainly to the extensive system of safety and control devices that every incinerator requires, regardless of size. This system is a significant part of the cost of incinerators and keeps the cost of increasing size relatively small in the small size range.

Figures 186 and 187 show that, as expected, the installed system cost as well as the incinerator cost for catalytic units are greater than for the same size thermal units.

However, when heat exchangers are added, this relationship is not apparent. Figures 188 and 189 show that catalytic units with heat exchange may tend to be slightly more expensive than a similar thermal unit; but there is not enough data to show a definite separation, and one curve can represent both types of units.

The savings in operating cost realized by heat recovery are illustrated in Figures 190 and 191. In each case, the cost of operating without heat recovery is two to three times that of operating with heat recovery. But it is apparent that the lines converge as the flow rate decreases, and that at some point heat recovery will not be advantageous.

Figures 192 and 193 show this same basic relationship when the annual capital charges are considered along with the direct operating costs. Due to the higher capital charges of units with heat exchange, the curves are closer and will converge at a higher flow rate than in Figures 190 and 191. Where these curves converge, heat recovery loses its economic advantage.

The economic relationship of catalytic incinerators and thermal units is illustrated by Figure 194. Even when the catalytic units' higher capital charges are considered, the savings in fuel keep the total cost of catalytic units lower than that of similar thermal units.

Electrostatic Precipitators

Electrostatic precipitators were used in only two applications in this report: vertical lime rock kilns and rotary lime sludge kilns. Figure 195 shows that the precipitator and installed system costs correlate well for both applications. The high efficiency unit costs 20 to 30% more than the medium efficiency unit, for the precipitator alone as well as for the installed system. The cost of the installed system was from about 2 to 2.5 times that of the precipitator alone.

(Text continued on page 696)

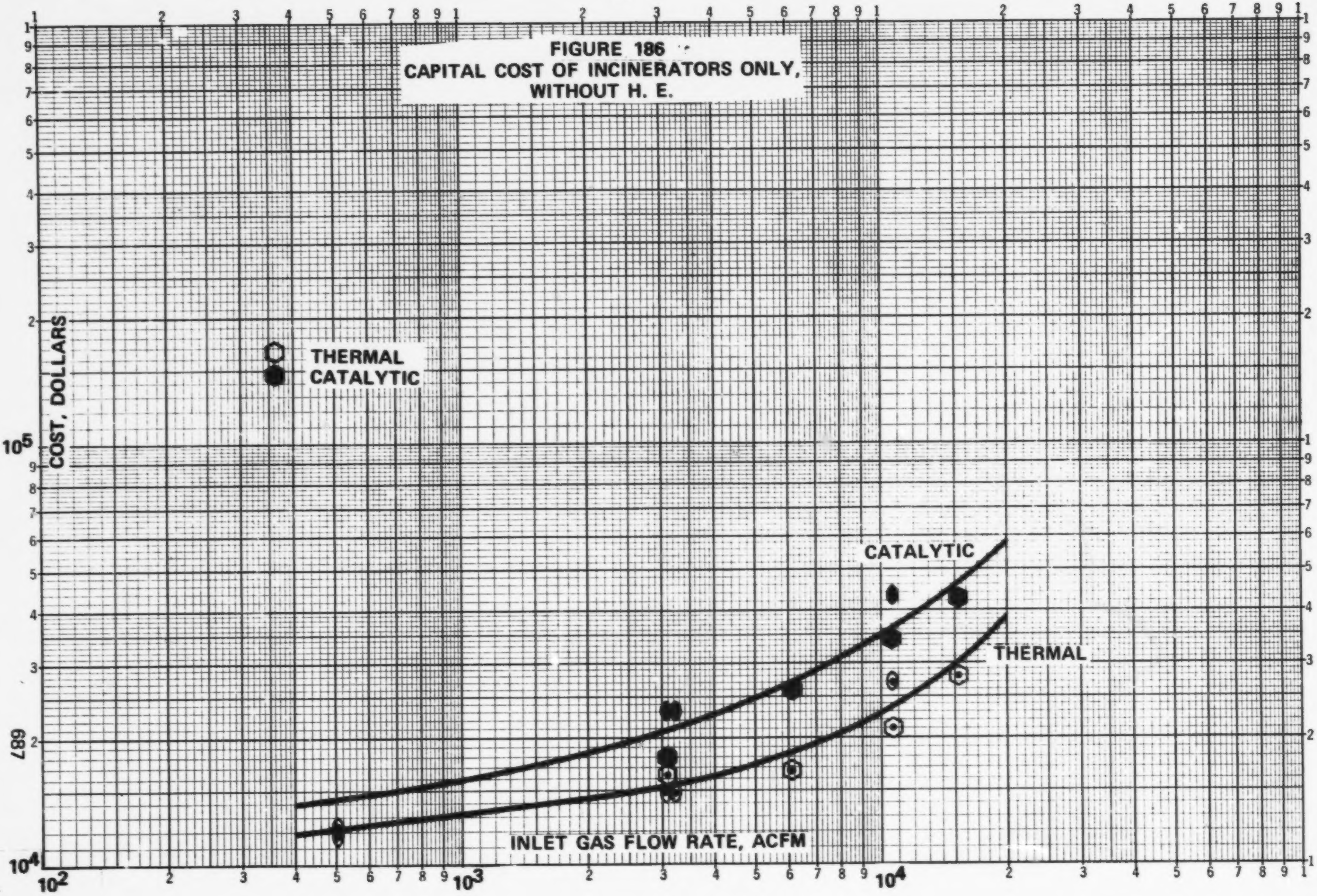


FIGURE 186
CAPITAL COST OF INCINERATORS ONLY, WITHOUT H. E.

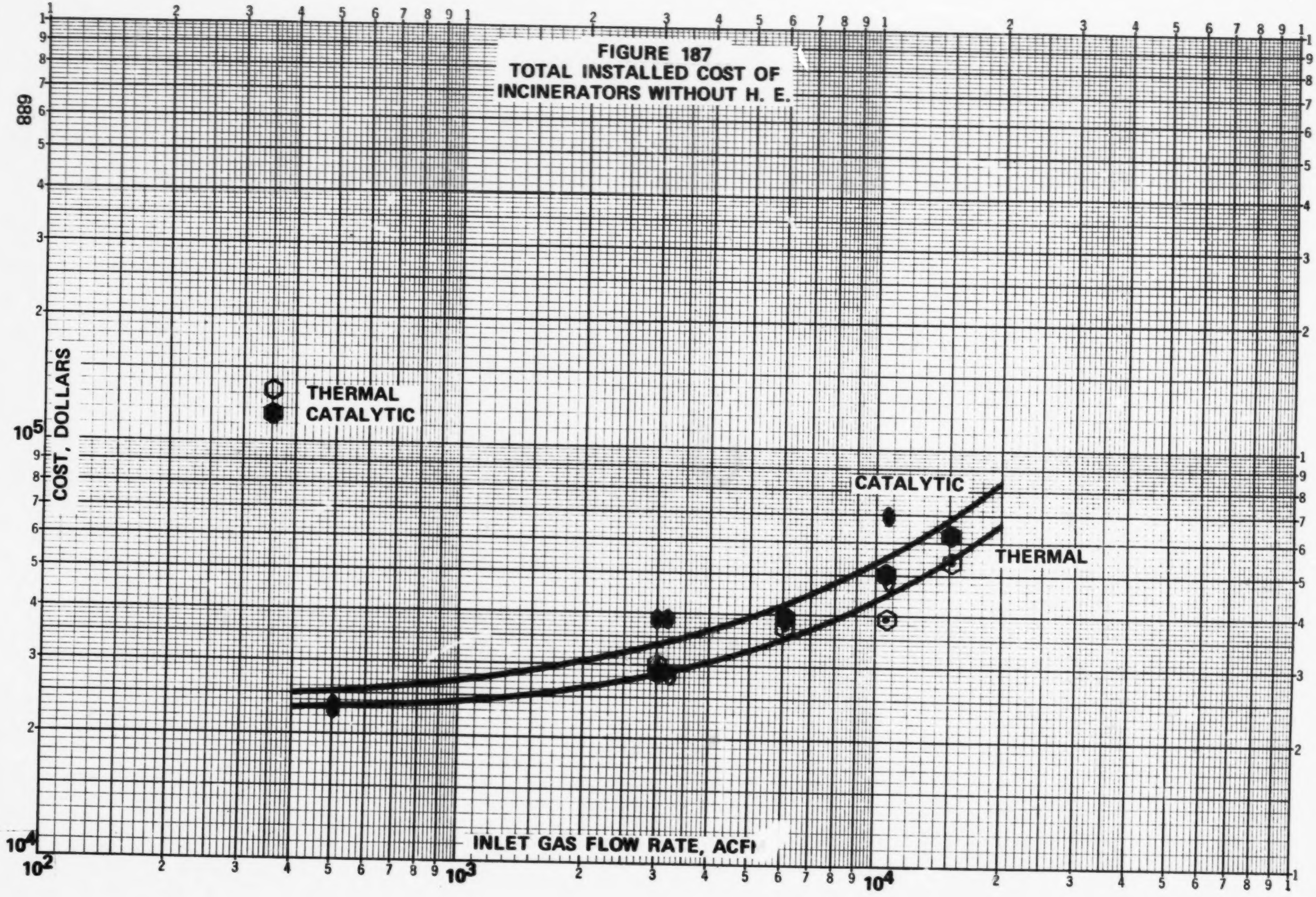
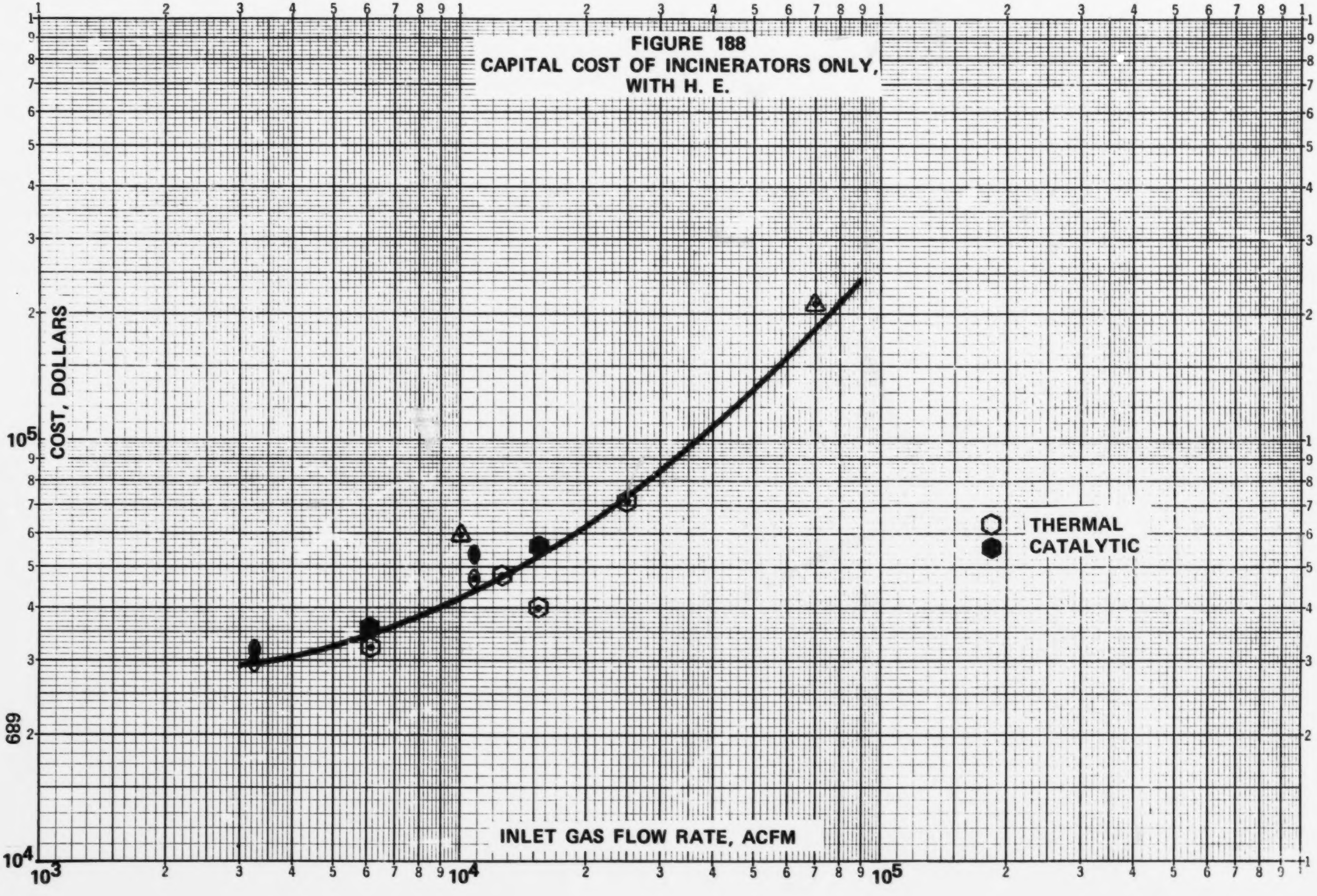


FIGURE 187
TOTAL INSTALLED COST OF
INCINERATORS WITHOUT H. E.

FIGURE 188
CAPITAL COST OF INCINERATORS ONLY, WITH H. E.



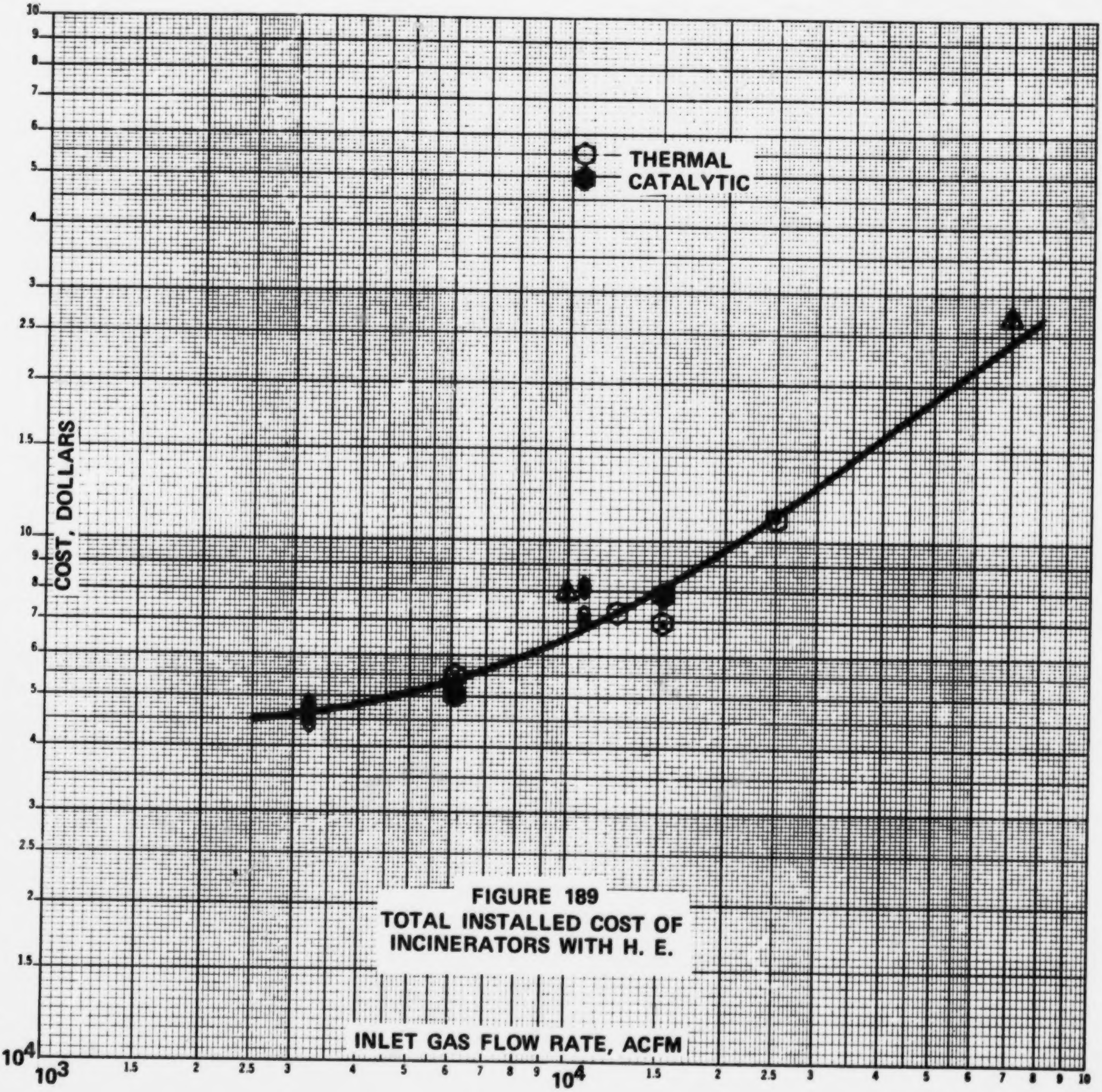


FIGURE 189
TOTAL INSTALLED COST OF
INCINERATORS WITH H. E.

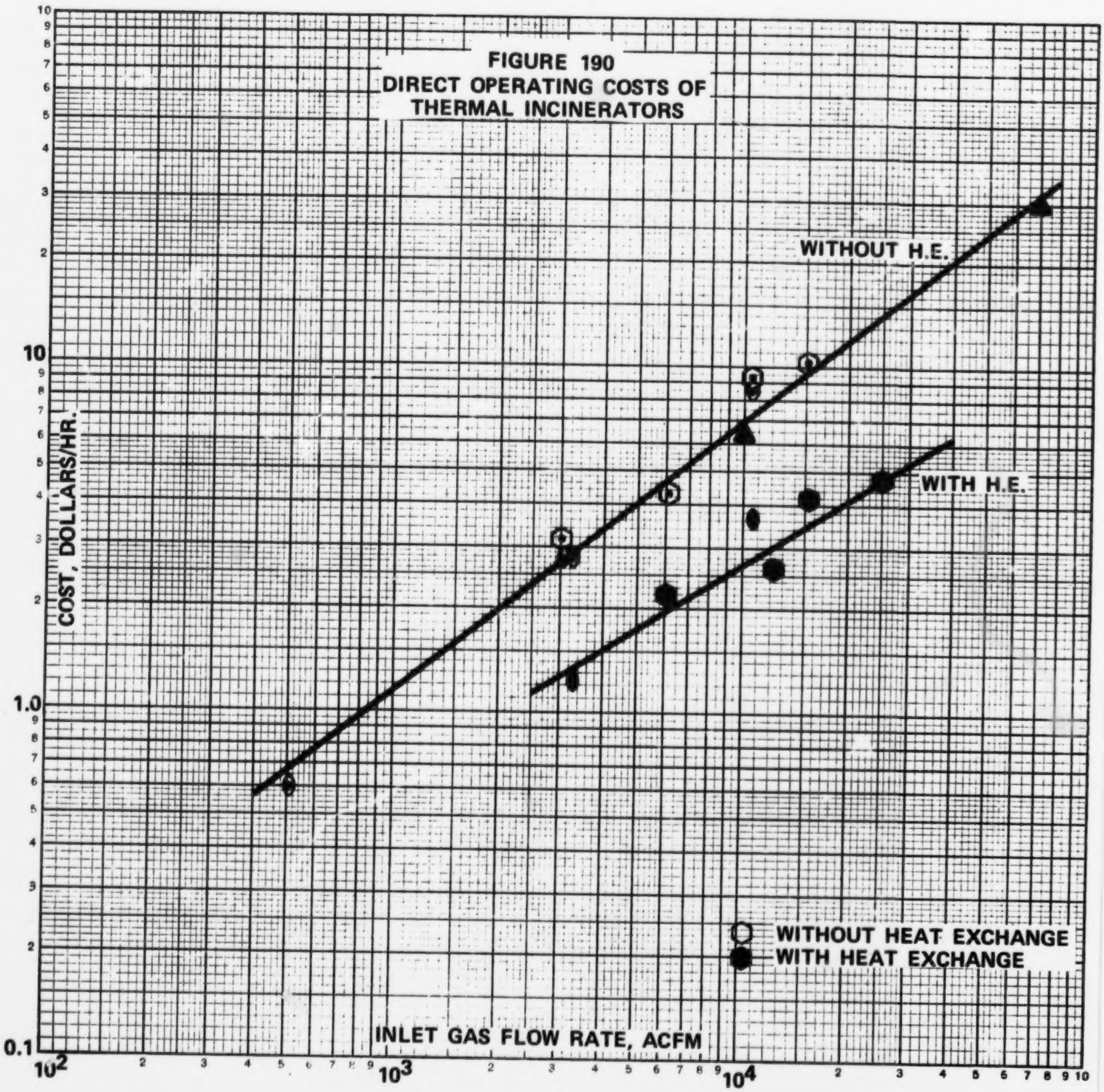


FIGURE 190
DIRECT OPERATING COSTS OF
THERMAL INCINERATORS
WITHOUT H.E.
WITH H.E.
COST, DOLLARS/HR.
10
1.0
0.1
INLET GAS FLOW RATE, ACFM
10^2
10^3
10^4
WITHOUT HEAT EXCHANGE
WITH HEAT EXCHANGE

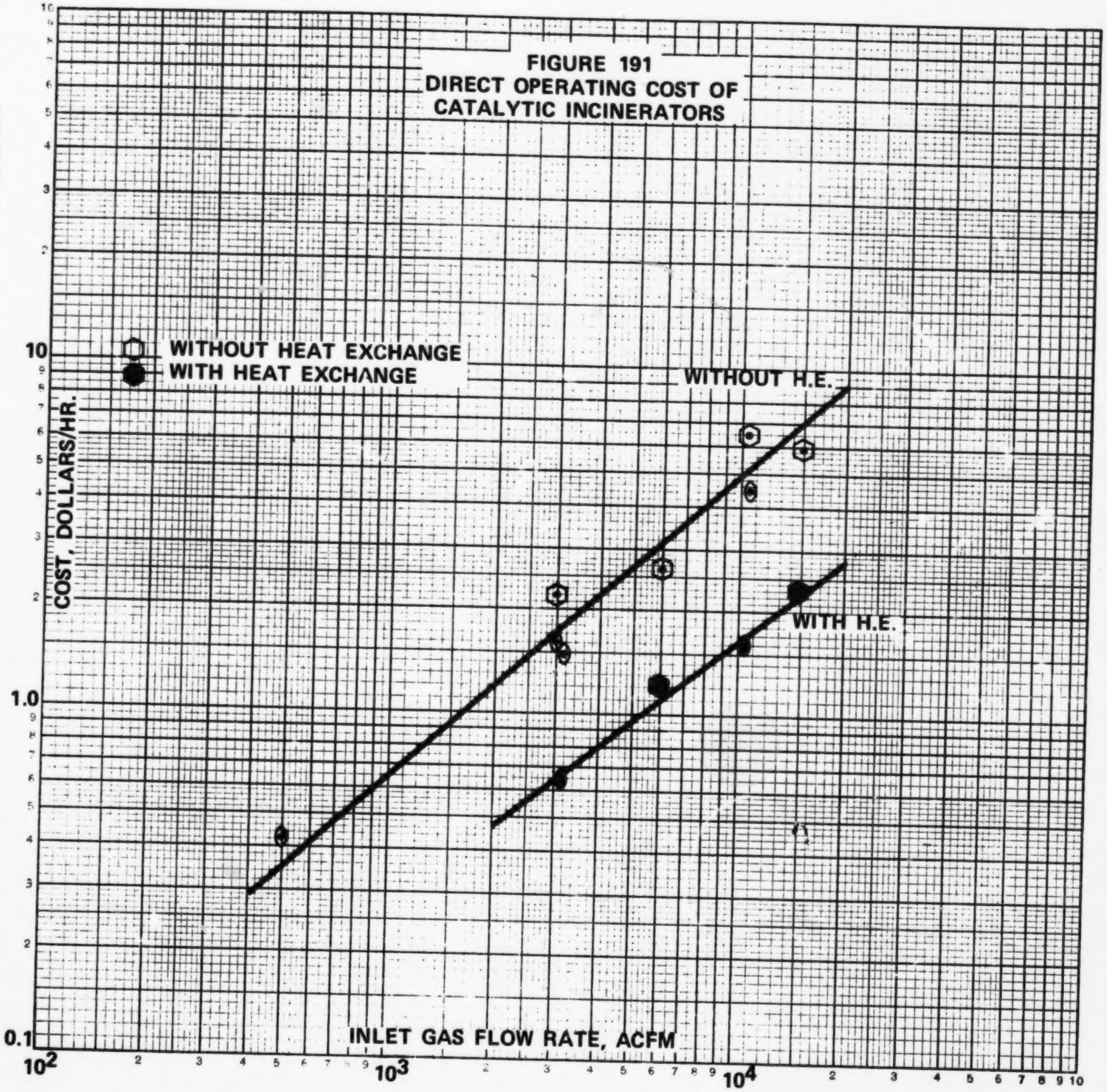


FIGURE 191
DIRECT OPERATING COST OF
CATALYTIC INCINERATORS

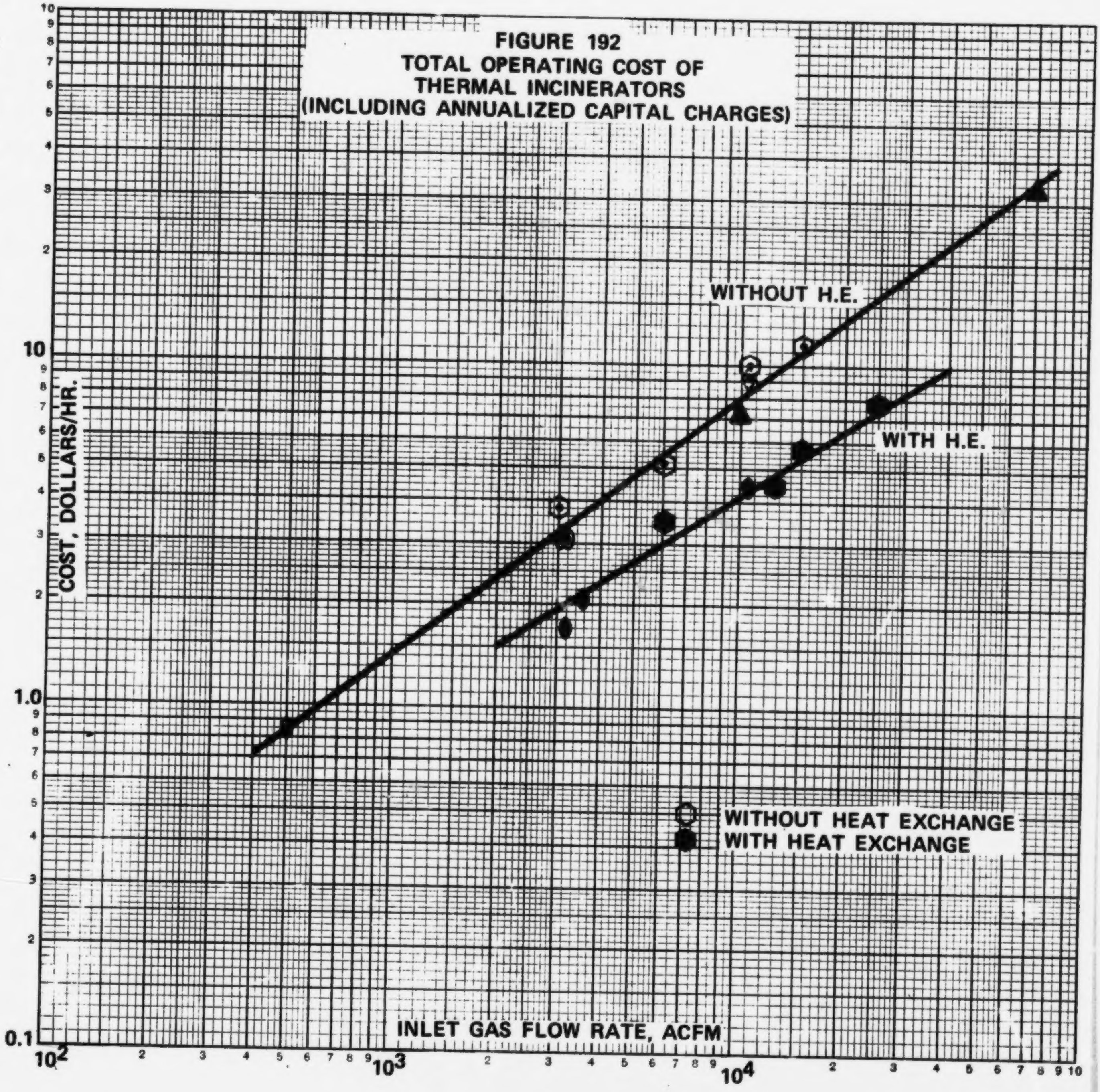


FIGURE 192
TOTAL OPERATING COST OF
THERMAL INCINERATORS
(INCLUDING ANNUALIZED CAPITAL CHARGES)

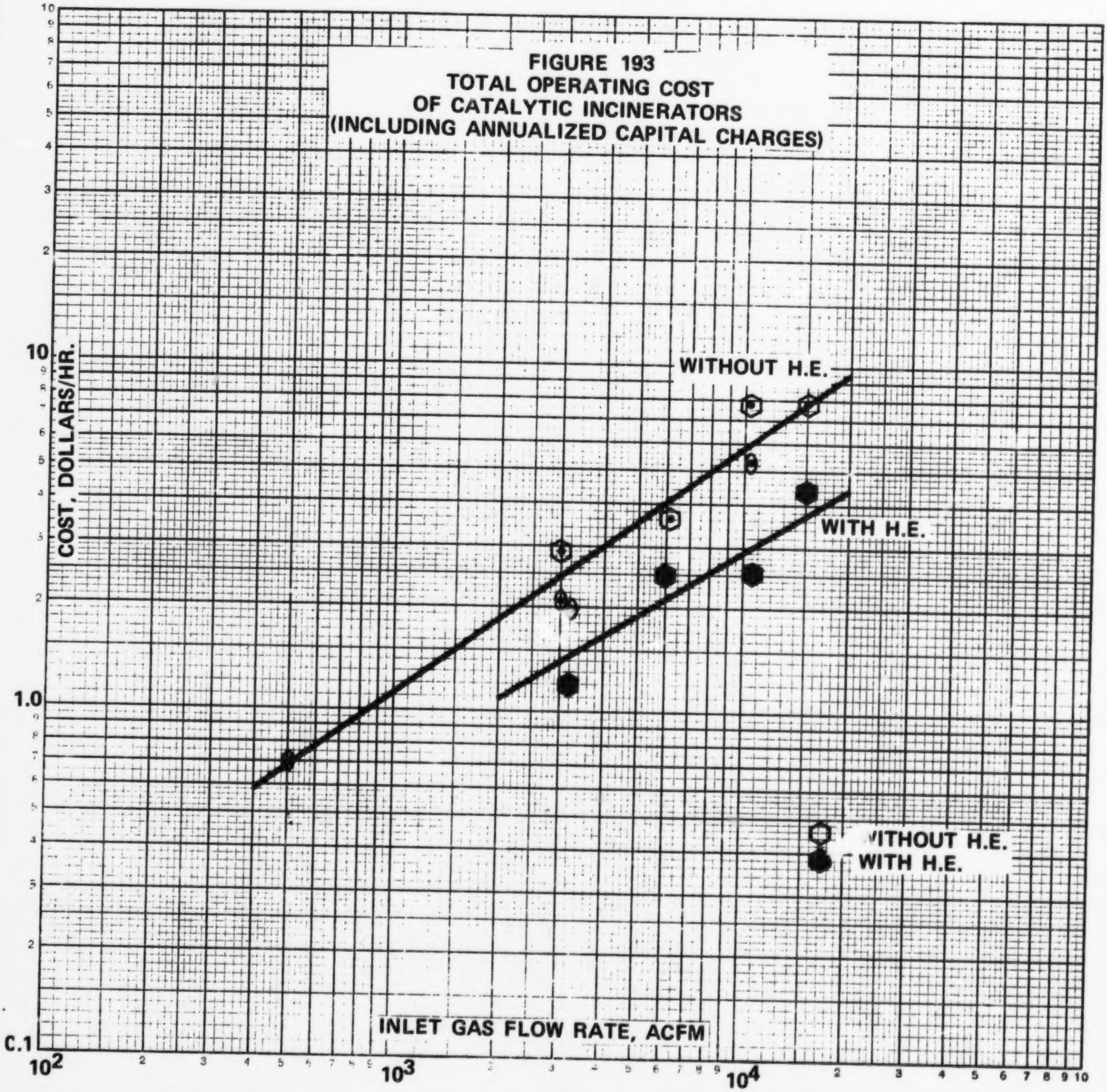


FIGURE 193
TOTAL OPERATING COST
OF CATALYTIC INCINERATORS
(INCLUDING ANNUALIZED CAPITAL CHARGES)

FIGURE 194
TOTAL OPERATING COST OF INCINERATORS WITHOUT H. E. (INCLUDING ANNUALIZED CAPITAL CHARGES)

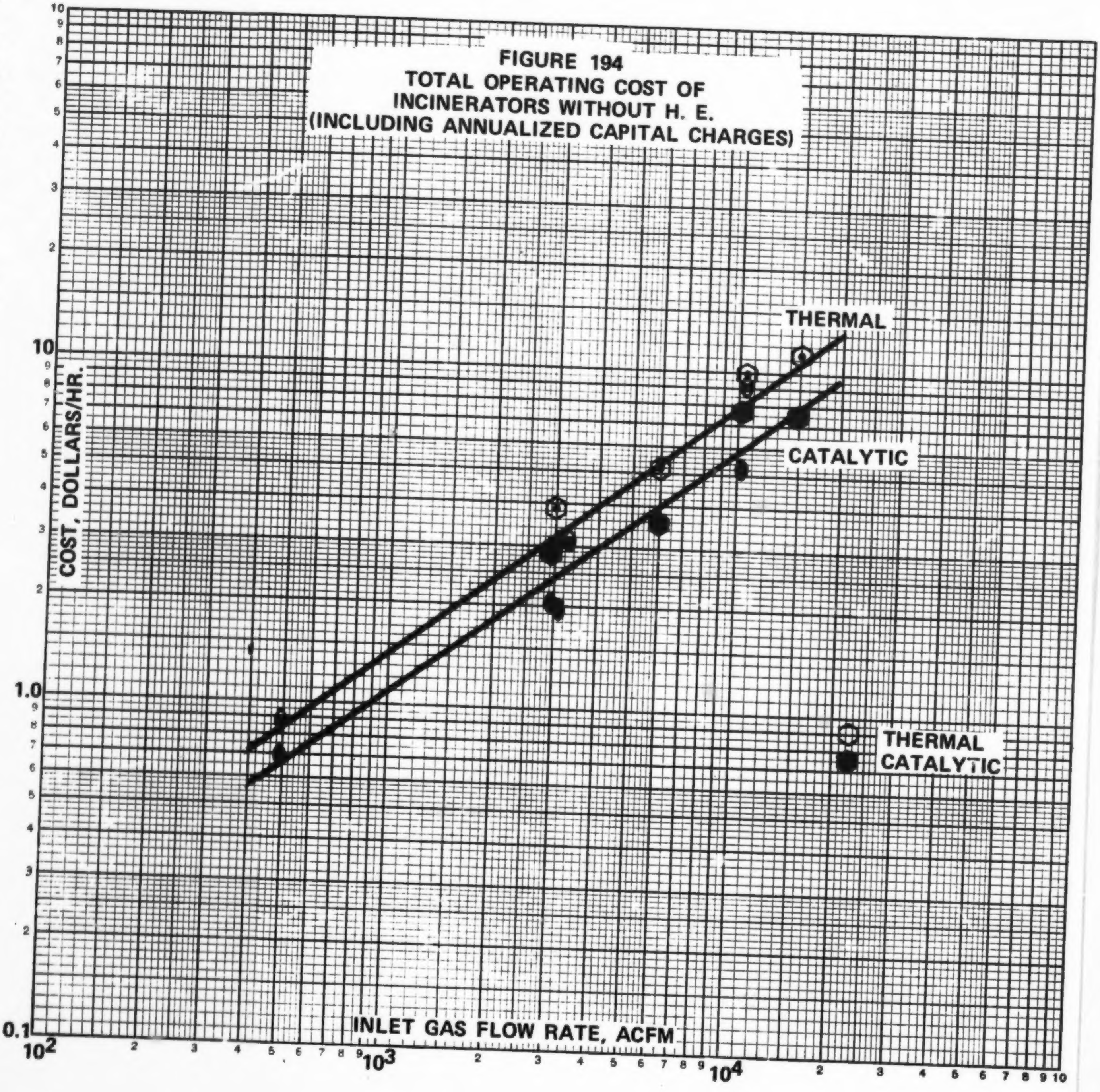


Figure 196 shows the direct operating costs of precipitators. These, too, correlated well. The data shows that the cost of operating a high efficiency unit is about 25 to 35% more than that of operating a medium efficiency unit. Nearly all of the increase is due to the increased use of electrical power.

Carbon Adsorption

Carbon adsorption was used in only one application in this report. The only correlations necessary were made when the costs were presented in the previous section of this report.

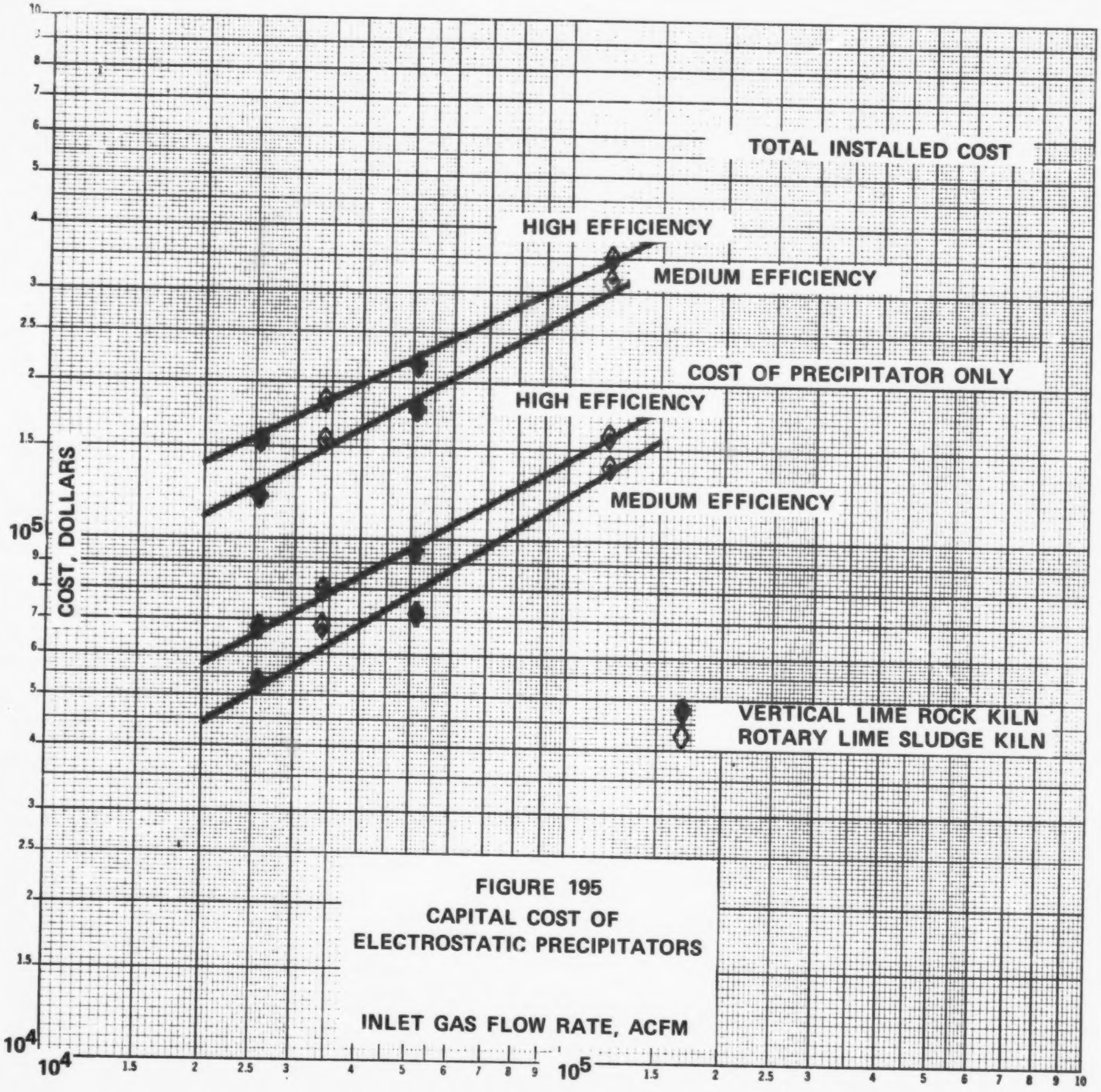


FIGURE 195
CAPITAL COST OF
ELECTROSTATIC PRECIPITATORS

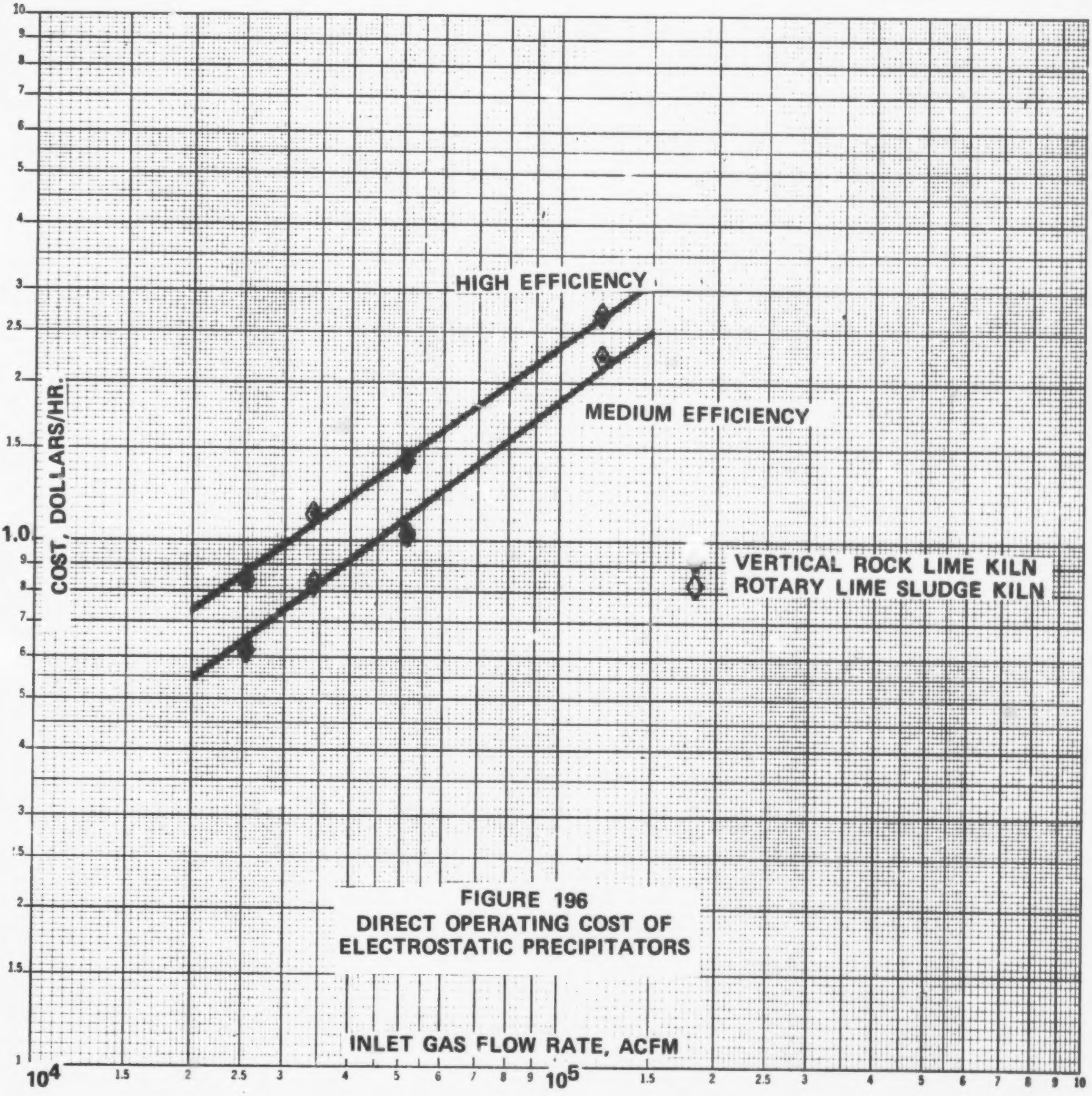


FIGURE 196
DIRECT OPERATING COST OF
ELECTROSTATIC PRECIPITATORS

APPENDIX I

SPECIFICATIONS FOR ABATEMENT EQUIPMENT

1. SCOPE

A. This specification covers vendor requirements for air pollution control equipment for the subject process. The intent of the specification is to describe the service as thoroughly as possible so as to secure vendor's proposal for equipment which is suitable in every respect for the service intended. Basic information is tabulated in sections 2 and 3. The vendor should specify any of the performance characteristics which cannot be guaranteed without samples of process effluent.

B. The vendor shall submit a bid showing three separate prices as described below.

1. All labor, materials, equipment, and services to furnish one pollution abatement device together with the following:

a. All ladders, platforms and other accessways to provide convenient access to all points requiring observation or maintenance.

b. Foundation bolts as required.

c. Six (6) sets of drawings, instructions, spare parts list, etc., pertinent to the above.

2. Auxiliaries including

(a) Fan(s)

(b) Pump(s)

(c) Damper(s)

(d) Conditioning Equipment

(e) Dust Disposal Equipment

3. A turnkey installation of the entire system including the following installation costs:

(a) Engineering

(b) Foundations & Support

(c) Ductwork

(d) Stack

(e) Electrical

- (f) Piping
- (g) Insulation
- (h) Painting
- (i) Startup
- (k) Performance Test
- (l) Other (including general tradework such as erection, rigging, etc.)

C. For the "pollution abatement device only" quotation, the vendor shall furnish the equipment FOB point of manufacture, and shall furnish as a part of this project competent supervision of the erection, which shall be by others.

D. Vendor shall furnish* the following drawings, etc., as a minimum:

1. With his proposal:
   - a. Plan and elevation showing general arrangement.
   - b. Typical details of collector internals proposed.
   - c. Data relating projected performance with respect to pressure drop, gas absorption efficiency and particulate removal efficiency to operating parameters such as gas flow.
2. Upon receipt of order:
   - a. Proposed schedule of design and delivery.
3. Within 60 days of order:
   - a. Complete drawings of equipment for approval by customer.
4. 30 days prior to shipment:
   - a. Certified drawings of equipment, six sets
   - b. Installation instructions, six sets
   - c. Starting and operating instructions, six sets
   - d. Maintenance instructions and recommended spare parts lists, six sets

E. The design and construction of the collector and auxiliaries shall conform to the general conditions given in Section 5, and to good engineering practice.

---

*This is a typical request. The member companies are NOT to furnish this material under the present project.

4. PROCESS PERFORMANCE GUARANTEE

A. The equipment will be guaranteed to reduce the particulate and/or gas contaminant loadings as indicated in the service description.

B. Performance test will be conducted in accordance with IGCI test methods where applicable.

C. Testing shall be conducted at a time mutually agreeable to the customer and the vendor.

D. The cost of the performance test is to be included in vendor's turnkey proposal.

E. In the event the equipment fails to comply with the guarantee at the specified design conditions, the vendor shall make every effort to correct any defect expeditiously at his own expense. Subsequent retesting to obtain a satisfactory result shall be at the vendor's expense.

5. GENERAL CONDITIONS

A. Materials and Workmanship

Only new materials of the best quality shall be used in the manufacture of items covered by this specification. Workmanship shall be of high quality and performed by competent workmen.

B. Equipment

Equipment not of vendor's manufacture furnished as a part of this collector shall be regarded in every respect as though it were of vendor's original manufacture.

C. Compliance with Applicable Work Standards and Codes

It shall be the responsibility of the vendor to design and manufacture the equipment specified in compliance with the practice specified by applicable codes.

D. Delivery Schedules

The vendor shall arrange delivery of equipment under this contract so as to provide for unloading at the job site within a time period specified by the customer. Vendor shall provide for expediting and following shipment of materials to the extent required to comply with delivery specified.

APPENDIX II

INSTRUCTIONS FOR SUBMITTING

COST DATA

Two forms (two copies each) are enclosed with each specification. These are for submitting:

(A) Estimated Capital Cost Data

(b) Annual Operating Cost Data

These forms will also be used to exhibit averages of the three cost estimates for each process and equipment type. Because your costs will be averaged with those of other IGCI members, it is necessary to prepare them in accordance with instructions given in the following paragraphs.

(A) Estimated Capital Cost Data

The upper part of this form should already be filled out for the particular application when you receive it. This information on operating conditions should be identical to that in the specification and is repeated here only for the convenience of those reading the form.

You should fill in the dollar amounts estimated in the appropriate spaces on the bottom half of the form. It should not be necessary to add any information other than the dollar amounts. If you wish to provide a description of the equipment proposed, please do so on one or more separate sheets of paper, and attach it to the form. If any item is not involved in the equipment you are proposing, please indicate this by writing "none" in the space rather than leaving it blank or using a zero.

(1) *The "gas cleaning device"* cost should be reported just as you would report a flange-to-flange equipment sale to the IGCI. That is, a complete device including necessary auxiliaries such as power supplies, mist eliminators, etc. Do NOT include such items as fans, solids handling equipment, etc., unless these are an *integral part of your gas cleaning device.*

(2) *"Auxiliaries"* are those items of equipment which are frequently supplied with the gas cleaning device. There is a purely arbitrary definition of those items included here and those included in the "Installation" Costs. Do NOT include any of the cost of erecting or installing auxiliaries in this category.

(3) *"Installation Cost"* should include all of the material not in (1) or (2) and the field labor required to complete a turnkey installation. In cases where the equipment supplier ordinarily erects the equipment but does not supply labor for foundations, etc., it is necessary to include an estimated cost for these items. General tradework, including rigging, erection, etc. should be included in the "Other" category.

The installation should be estimated for a new plant, or one in which there are no limitations imposed by the arrangement of existing equipment. Installation labor should be estimated on the basis that the erection will take place in an area where labor rates are near the U.S. average, and the distance from your plant is no more than 500 miles. Milwaukee, Wisconsin is an example of a city with near-average labor rates.

(B) Annual Operating Cost Data

Some of the information will be supplied by Air Resources, such as unit costs for labor and utilities, and annualized capital charges. You should fill in the usage figures for the complete abatement system in the units indicated below. Please include the unit price.

Labor	hrs/year
Maintenance Materials	Dollars/year
Replacement Parts	Dollars/year
Electric Power	kw-hr/year
Fuel	MMBTU/year
Water (Process)	MM gal/year
Water (Cooling)	MM gal/year
Chemicals	Dollars/year

Air Resources will average the consumption figures reported, and convert them to dollar values for inclusion in the final report.

Be sure that the operating factor, in hours per year, supplied by ARI, is used for estimating the utility and labor requirements.

APPENDIX III

CITY COST INDICES

Average 1969 Construction Cost & Labor Indices						Historical Average	
City	Index		City	Index		Year	Index
	Labor	Total		Labor	Total	1969	100
Albany, N.Y.	98	100	Milwaukee, Wi.	103	108	1968	91
Albuquerque, N.M.	86	95	Minneapolis, Mn.	99	98	1967	86
Amarillo, Tx.	87	84	Mobile, Al.	94	90	1966	83
Anchorage, Ak.	131	148	Montreal, Cn.	77	89	1965	79
Atlanta, Ga.	88	94	Nashville, Tn.	79	82	1964	78
Baltimore, Md.	90	93	Newark, N.J.	122	109	1963	76
Baton Rouge, La.	83	88	New Haven, Ct.	102	100	1962	74
Birmingham, Al.	79	86	New Orleans, La.	89	95	1961	72
Boston, Ma.	106	103	New York, N.Y.	132	118	1960	71
Bridgeport, Ct.	104	102	Norfolk, Va.	73	77	1959	69
Buffalo, N.Y.	104	107	OklahomaCity,Ok.	82	88	1958	67
Burlington, Vt.	86	90	Omaha, Nb.	90	93	1957	65
Charlotte, N.C.	70	75	Philadelphia, Pa.	106	101	1956	63
Chattanooga, Tn.	81	84	Phoenix, Az.	101	97	1955	59
Chicago, Ill.	107	103	Pittsburgh, Pa.	110	106	1954	58
Cincinnati, Oh.	108	104	Portland, Me.	82	87	1953	57
Cleveland, Oh.	121	112	Portland, Or.	102	103	1952	55
Columbus, Oh.	106	99	Providence, R.I.	98	97	1951	53
Dallas, Tx.	86	89	Richmond, Va.	76	79	1950	49
Dayton, Oh.	100	103	Rochester, N.Y.	110	107	1949	48
Denver, Co.	94	91	Rockford, Ill.	109	109	1948	48
Des Moines, Ia.	93	96	Sacramento, Ca.	117	110	1947	43
Detroit, Mi.	117	111	St. Louis, Mo.	110	103	1946	35
Edmonton, Cn.	80	83	Salt Lake City, Ut.	93	95	1945	30
El Paso, Tx.	77	83	San Antonio, Tx.	82	82	1344	29
Erie, Pa.	98	99	San Diego, Ca.	111	107	1943	29
Evansville, In.	93	97	San Francisco, Ca.	124	109	1942	28
Grand Rapids, Mi.	103	99	Savannah, Ga.	72	77	1941	25
Harrisburg, Pa.	90	92	Scranton, Pa.	94	96	1940	24
Hartford, Ct.	104	100	Seattle, Wa.	104	99	1939	23
Honolulu, Hi.	99	109	Shreveport, La.	82	89	1938	23
Houston, Tx.	92	89	South Bend, In.	99	97	1937	23
Indianapolis, In.	97	98	Spokane, Wa.	101	100	1936	20
Jackson, Ms.	73	75	Springfield, Ma.	99	97	1935	20
Jacksonville, Fl.	78	79	Syracuse, N.Y.	105	103	1934	20
Kansas City, Mo.	94	93	Tampa, Fl.	81	84	1933	18
Knoxville, Tn.	82	82	Toledo, Oh.	105	105	1932	17
Las Vegas, Nv.	115	107	Toronto, Cn.	84	93	1931	20
Little Rock, Ar.	78	81	Trenton, N.J.	114	103	1930	22
Los Angeles, Ca.	113	102	Tulsa, Ok.	85	89	1929	23
Louisville, Ky.	92	93	Vancouver, Cn.	81	91	1928	23
Madison, Wi.	95	98	Washington, D.C.	98	94	1927	23
Manchester,N.H.	89	92	Wichita, Ks.	85	90	1926	23
Memphis, Tn.	83	82	Winnipeg, Cn.	62	82	1925	23
Miami, Fl.	98	94	Youngstown, Oh.	107	106	1924	23

APPENDIX III-a

AVERAGE HOURLY LABOR RATES BY TRADE

Trade	1970	1969	1968	1967	1966
Common Building Labor	$5.00	$4.55	$4.10	$3.85	$3.65
Skilled Average	6.85	6.05	5.50	5.15	4.90
Helpers Average	5.15	4.65	4.20	4.00	3.85
Foremen (usually 35¢ over trade)	7.20	6.40	5.85	5.50	5.25
Bricklayers	7.15	6.40	5.85	5.55	5.35
Bricklayers Helpers	5.20	4.70	4.30	4.05	3.95
Carpenters	6.95	6.15	5.40	5.10	4.90
Cement Finishers	6.75	5.90	5.30	5.05	4.85
Electricians	7.50	6.45	5.95	5.60	5.45
Glaziers	6.25	5.50	5.10	4.75	4.60
Hoist Engineers	7.05	5.90	5.40	5.10	4.85
Lathers	6.60	5.95	5.45	5.20	5.05
Marble & Terrazzo Workers	6.45	5.60	5.25	5.05	4.90
Painters, Ordinary	6.20	5.45	5.05	4.75	4.50
Painters, Structural Steel	6.50	5.80	5.30	4.95	4.80
Paperhangers	6.30	5.60	5.15	4.75	4.55
Plasterers	6.60	5.95	5.50	5.15	5.00
Plasterers Helpers	5.30	4.85	4.45	4.15	4.00
Plumbers	7.75	6.90	6.15	5.75	5.55
Power Shovel or Crane Operator	7.20	6.20	5.65	5.35	5.05
Rodmen (Reinforcing)	7.30	6.35	5.80	5.45	5.15
Roofers, Composition	6.30	5.55	5.05	4.75	4.65
Roofers, Tile & Slate	6.35	5.60	5.10	4.85	4.80
Roofers Helpers (Composition)	4.75	4.45	4.00	3.75	3.55
Steamfitters	7.70	6.90	6.10	5.70	5.50
Sprinkler Installers	7.70	6.90	6.10	5.70	5.50
Structural Steel Workers	7.45	6.45	5.90	5.55	5.25
Tile Layers (Floor)	6.50	5.60	5.20	4.90	4.80
Tile Layers Helpers	5.25	4.80	4.35	4.15	4.05
Truck Drivers	5.15	4.60	4.30	3.95	3.65
Welders, Structural Steel	7.15	6.35	5.80	5.45	5.10

APPENDIX IV

STATISTICAL BASIS FOR DATA PRESENTATION

The cost quotations received from member companies have in every case been averaged and the resulting values presented graphically in the body of the report. Provided there is no more than a reasonable spread between the quotations, it is helpful to treat the data received as a random selection from among a "population" of twenty or so potential bidders. Statistical values for the confidence limits of the mean cost have been calculated.

Calculation Method – The calculations performed by ***CONLIM are based on the following formulas:

$$\text{Confidence limits} = \bar{X} \pm s t_{n-1;\gamma}$$

where

$\bar{X}$ = the sample mean, based on three bids in most cases

s = the sample standard deviation

$t_{n-1;\gamma}$ - the (γ X 100) percentage point of the student-t distribution with n-1 degrees of freedom

Size of sample - n, usually three

$$\text{Sample mean value} = \frac{1}{n} \sum_{i=1}^{n} X_i$$

$$\text{Variance of sample} = \frac{1}{n} \sum_{i=1}^{n} (X_i - \bar{X})^2$$

$$\text{Standard deviation of sample} = \sqrt{\frac{\Sigma (X_i - \bar{X})^2}{n}}$$

$$\text{Estimated population standard deviation} = \sqrt{\frac{\Sigma (X_i - \bar{X})^2}{n-1}}$$

$$\text{Standard error of mean} = \frac{S}{\sqrt{n}}$$

where

S - sample standard deviation

When the population is finite, a correction factor of $(\frac{N-n}{N})$ is included in both the variance and the standard deviation computations, as follows:

$$S_f^2 = S^2(\frac{N-n}{N})$$

where

S_f^2 = the corrected variance for finite populations

S^2 - the non-corrected variance for infinite populations

N - the population size, usually taken as 20

n - the sample size, usually three

The results are presented graphically using a solid line on log-log paper for the mean cost vs. equipment size, and dotted lines for the 75% and 90% confidence intervals based on three bids (or the actual number of bids received) out to an approximate population of 20 possible bidders.

APPENDIX V

LIST OF STANDARD ABBREVIATIONS

feet or foot	ft
inch or inches	in.
ton or tons	ton
pound or pounds	lb
hours or hours	hr
minute or minutes	min
parts per million	ppm
grain or grains	gr
weight percent	wt.%
actual cubic feet per minute	ACFM
standard cubic feet per minute	SCFM
dry standard cubic feet per minute	DSCFM
standard cubic feet	SCF
actual cubic feet	ACF
British thermal units	Btu
odor units	o.u.
volume	vol
mole	mol
gallon	gal
per cent	%
dollars	$
degrees Fahrenheit	°F
pounds per square inch gauge	psig
change of pressure (delta pressure)	ΔP
water column (pressure)	w.c.
change of temperature (delta temperature)	ΔT
temperature	Temp
feet per minute	FPM
dry standard cubic feet	DSCF
cubic feet	ft^3
revolutions per minute	rpm
gallons per minute	gpm
millions (10^6)	MM
atmospheres gage (pressure)	atmg
milligrams	mg
micrograms	yg
international unit	IU
hundred weight (100 pounds)	cwt
hydrocarbon	Hcbn
United States pharmacopoeia	USP